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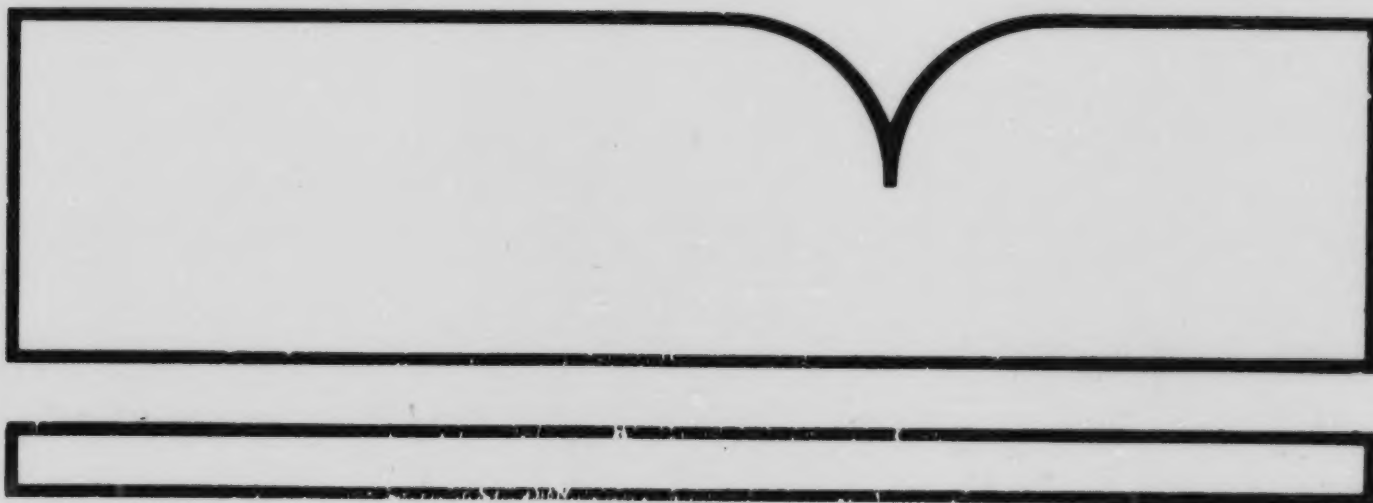
On-Farm Improvements to Reduce Sediment and
Nutrients in Irrigation Return Flow

Washington State Univ., Pullman

Prepared for

Robert S. Kerr Environmental Research Lab.
Ada, OK

Feb 84



U.S. Department of Commerce
National Technical Information Service

NTIS

PB84-159243

Evaluation and Demonstration of the
Chemically Active Fluid Bed

GCA Corp., Chapel Hill, NC

Prepared for

Industrial Environmental Research Lab.
Research Triangle Park, NC

Feb 84

PB84-159243

EPA-600/7-84-031
February 1984

EVALUATION AND DEMONSTRATION OF THE
CHEMICALLY ACTIVE FLUID BED

Final Report

by

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EPA Contract No. 68-02-3168
Work Assignment No. 86

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U.S. ENVIRONMENTAL PROTECTION AGENCY
RESEARCH TRIANGLE PARK, NC 27711

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SPRINGFIELD, VA. 22161

TECHNICAL REPORT DATA (Please read Instructions on the reverse before completing)			
1. REPORT NO. EPA-600/7-84-031		3. RECIPIENT'S ACCESSION NO. PB8 4 15924 3	
4. TITLE AND SUBTITLE Evaluation and Demonstration of the Chemically Active Fluid Bed		5. REPORT DATE February 1984	
		6. PERFORMING ORGANIZATION CODE	
7. AUTHOR(S) Rebecca E. Sommer, Arthur S. Werner, and Z. Kowszun*		8. PERFORMING ORGANIZATION REPORT NO. GCA-TR-CH-83-04	
9. PERFORMING ORGANIZATION NAME AND ADDRESS GCA/Technology Division Chapel Hill, North Carolina 27514		10. PROGRAM ELEMENT NO.	
		11. CONTRACT/GRANT NO. 68-02-3168, Task 86	
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Task Final; 5/75 - 7/81	
		14. SPONSORING AGENCY CODE EPA/600/13	
15. SUPPLEMENTARY NOTES IERL-RTP project officer is Samuel L. Rakes, Mail Drop 4; 919/541-2828. Coauthor Kowszun is with Esso Research Centre, Abingdon, Oxfordshire, OX136AE, England.			
16. ABSTRACT The report gives results of the operation of a 17-MW Chemically Active Fluid Bed (CAFB) demonstration unit, retrofitted to a natural gas boiler. The CAFB process gasifies high-sulfur, high-metals-content liquid and solid fuels. Residual oil, lignite, and bituminous coal were gasified separately or together between November 1979 and June 1981. Oil was gasified for 540 hours. A maximum continuous-rating test sustained 22 MW of electrical energy for 2 hours. Sulfur removal efficiencies of up to 90% were achieved during oil gasification. Lignite and bituminous coal were gasified for 42 hours: sulfur removal efficiencies of up to 70% were obtained. Design and operational areas where upgrading would be beneficial were identified. Continuous monitors were used to measure boiler flue gas emissions of SO ₂ , NO _x , CO, oxygen, CO ₂ , and opacity. Periodic manual emission tests were conducted for particulate, SO ₂ , and NO _x , using EPA reference methods. Emissions of these three criteria pollutants were generally lower than New Source Performance Standards for utility boilers, although occasionally excessive particulate and SO ₂ emissions were observed. NO _x emissions were consistently lower than those from natural gas combustion. Results of detailed chemical analyses and biological assays are reported.			
17. KEY WORDS AND DOCUMENT ANALYSIS			
a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution	Gasification	Pollution Control	13B
Fluidized Bed Processing		Stationary Sources	13H, 07A
Residual Oils	Desulfurization	Chemically Active Fluid	11H, 21D 07D
Lignite	Particles	Beds	08G 14G
Bituminous Coal	Dust	Particulate	11G
Coal Gasification	Nitrogen Oxides		07B
Flue Gases	Fly Ash		21B
18. DISTRIBUTION STATEMENT Release to Public		19. SECURITY CLASS (This Report) Unclassified	21. NO. OF PAGES 222
		20. SECURITY CLASS (This page) Unclassified	22. PRICE

DISCLAIMER

This Final Report was furnished to the Environmental Protection Agency by GCA Corporation, GCA/Technology Division, 213 Burlington Road, Bedford, Massachusetts 01730, in fulfillment of Contract No. 68-02-3168, Task No. 86. The opinions, findings, and conclusions expressed are those of the authors and not necessarily those of the Environmental Protection Agency. Mention of company or product name is not to be considered as an endorsement by the Environmental Protection Agency.

ABSTRACT

Results are reported on the operation of a 17 MW Chemically Active Fluid Bed (CAFB) demonstration unit retrofit to an existing natural gas boiler. The CAFB process is designed to gasify high sulfur, high metals content liquid and solid fuel. Residual oil, lignite and bituminous coal were gasified separately or together during the period from November 1979 through June 1981. Oil was gasified for 540 hours. A maximum continuous rating test sustained 22 MW of electrical energy for two hours. Sulfur removal efficiencies of up to 90 percent were achieved during oil gasification. Lignite and bituminous coal were gasified for a total of 42 hours during which time sulfur removal efficiencies of up to 70 percent were obtained. Design and operational areas which need to be upgraded were identified. Continuous monitors were used to measure boiler flue gas emissions of sulfur dioxide, nitrogen oxides, carbon monoxide, oxygen, carbon dioxide and opacity. Periodic manual emission tests were conducted for particulate, sulfur dioxide and nitrogen oxides using EPA Reference Methods. Emissions of these three criteria pollutants were generally lower than New Source Performance Standards for utility boilers, although occasionally excessive particulate and sulfur dioxide emissions were observed. Nitrogen oxide emissions were consistently lower than emissions from natural gas combustion. Potential multi-media impacts from the three major discharge streams from the CAFB - boiler flue gas, spent solids from the gasifier bed, and sulfur recovery system ash - were characterized using the Phased Approach to Environmental Assessments developed by EPA's Industrial Environmental Research Laboratory. Results of detailed chemical analyses and biological assays are reported.

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ACKNOWLEDGEMENTS

The authors gratefully acknowledge the guidance and technical assistance provided by the EPA Project Officer, Mr. Samuel Rakes. We also thank Mr. Richard McMillan of Foster Wheeler Energy Corporation for assembling the operating data and providing insightful counsel regarding interpretation of the data.

SECTION 1

EXECUTIVE SUMMARY

BACKGROUND

The Chemically Active Fluid Bed (CAFB) process was developed in the late 1960's by the Esso Research Center, Abingdon (ERCA), England as a means to generate electrical energy from high sulfur, high metal heavy fuel oil.¹ In the CAFB process oil or coal is fed continuously into a fluidized bed of lime maintained at 870°C (1600°F). Air is preheated, and fed to the gasifier in substoichiometric proportions. The fuel entering the gasifier is vaporized, oxidized, and cracked to produce a low-Btu, low-sulfur gas which after passing through cyclones to remove particulate matter, is burned in a conventional gas boiler. Sulfur contained in the oil initially forms various gaseous compounds which then react with the bed lime to yield solid calcium sulfide. Some of the fuel bound trace elements are also bound to the bed stone. The sulfided lime is cycled to a regeneration unit where it is oxidized to produce calcium oxide which is returned to the gasifier and sulfur dioxide which is sent to a sulfur recovery unit.

During the early and mid 1970's, the process was demonstrated at a 2.9 MW pilot plant to provide energy efficiently from both liquid and solid fuels while simultaneously limiting emissions of sulfur dioxide, nitrogen oxides, vanadium, and nickel. Based on the successful pilot plant tests in England, the U.S. Environmental Protection Agency (EPA) and the Texas based Central Power and Light Company (CP&L) commissioned Foster Wheeler Energy Corporation (FWEC) to construct a 17 MW demonstration plant in San Benito, Texas. The agreement with CP&L, signed in May 1975, provided for construction and operation of the demonstration plant. The contract with EPA (Contract No. 68-02-2106) covered engineering design, labor for start-up, fuels and feedstocks, and environmental monitoring. Design of the unit, which was based on the ERCA design and operating experience, was initiated in January 1975 and construction was completed in January 1979. The plant incorporated FWEC's proprietary sulfur recovery system - RESOXTM. Over 130 contractors

and subcontractors were employed during construction. A test program using heavy oil, lignite and bituminous coal was conducted between October 1979 and June 1981.

This report describes the design, operation and environmental aspects of the San Benito demonstration plant.

PLANT OPERATION

Oil, lignite and bituminous coal were gasified separately or together in the CAFB unit, which was retrofit on the front end of an existing natural gas boiler, during the period from November 1979 through June 1981. Eleven distinct test runs were conducted over this time span. Operating conditions and fuels employed for the runs are summarized in Table 1-1. Residual oil was gasified during the first four runs which totaled 343.5 hours of operation. The fifth gasification run was the longest and most successful. All three systems, the gasifier, regenerator and RESOXTM reactor, were operational during the run. During the fifth run oil was the primary feedstock as it was gasified for 196 hours, however, lignite was introduced and gasified in combination with the oil for the final three hours of operation. During the second run a maximum continuous rating test sustained 22 MW of electrical energy on oil for a two hour period. Problems with the transfer of bed stone between the gasifier and regenerator limited regenerator operation during oil gasification to a total of 213 hours. The RESOXTM sulfur recovery system was operated during oil gasification for a total of 115 hours. Sulfur removal efficiencies of up to 90 percent were obtained.

West Texas lignite was gasified for 10.5 hours but problems with the coal feed system precluded additional operation. An eastern bituminous coal was gasified for a total of 31.7 hours during the final four runs. Again coal feed problems caused premature termination of two of the runs. Throughout the final two runs the boiler flames were strong and well-defined. A sulfur removal efficiency of 70 percent was achieved during the final run.

Areas in which improvements are necessary include:

TABLE 1. SUMMARY OF THE 11 GASIFICATION RUNS BETWEEN DECEMBER 1979 and JUNE 1981

Run	Dates	Hours of gasification	Fuels used	Regenerator operation	RESOX TM operation	Boiler load/ product gas	Comments
1	12/19/79- 12/20/79	32	Oil	No	No	4 MW - 6 hrs 9-10 MW - 24 hrs	
2	1/27/80- 1/31/80	77.5	Oil	No	No	9-11 MW - 48 hrs 14-15 MW - 8 hrs 20-22 MW - 3 hrs	Maintained 22 MW for maximum continuous rating.
3	4/8/80- 4/14/80	144	Oil	No	No	7.2-15 MW	
4	7/29/80- 8/1/80	90	Oil	Yes	Yes 17 hrs	10.5 MW 72 hrs product/natural gas	First use of N ₂ for material transfer between gasifiers and regenerator.
5	9/24/80- 10/1/80	196	Oil Lignite, 2 hrs	Yes	Yes	8.5 MW - product gas 2-hrs coal provided 30% of 8.5 MW	During this run GCA performed environmental assessment
6	11/20/80- 11/25/80	79	Oil Lignite Oil/Lignite	No	No	5-6 MW	
7	12/9/80 12/12/80	5.5	Oil Oil/Lignite Lignite	Yes	No	9 MW 7.5 MW	The gasifier bed appeared to be releasing SO ₂ .
8	2/81	4	Lignite	Yes	No		Run aborted.
9	3/11/83- 3/12/83	10	Oil, 3 hrs Oil/coal, 3 hrs Coal, 4 hrs	No	No	8.7 MW	Approximately 4.5 MW was from product gas.
10	6/5 - 6/6/81 6/7 - 6/8/81 6/9 - 6/10/81	19.5 17.5 19	Oil Oil, Oil/coal Oil, Oil/coal	Yes	No	9.0 MW	
11	6/22/81 - 6/23/81	16	Oil, 3 hrs Oil/coal, 5 hrs Coal, 8 hrs	No	No	4.8 MW - product gas 5.9 MW - natural gas	The bed from run 10 was reused in this run.

- More efficient clean up of product gas entering the boiler. The single stage cyclones employed at San Benito were, at times, inadequate.
- A sturdier, more reliable coal feed system. The coal feed system was very sensitive to coal physical properties and was prone to plugging under some conditions.
- A burn-up cell for the carbon on the cyclone fines would be desirable.
- More reliable rotary seals, valves, dampers, augers, and filters are needed to better withstand the severe conditions of CAFB operation.

ENVIRONMENTAL ASPECTS

The demonstration unit was evaluated to determine its capability to control flue gas criteria pollutant emissions and to estimate potential multi-media pollution impacts. Continuous monitors were used by FWEC to measure sulfur dioxide, nitrogen oxides, carbon monoxide, oxygen, carbon dioxide and opacity. In addition, periodic manual emission tests were conducted by FWEC for particulate, sulfur dioxide and nitrogen oxides using EPA reference methods 5, 6 and 7, respectively. Potential multi-media environmental impacts were determined under a separate contract conducted by GCA/Technology Division (EPA Contract no. 68-02-2695). This program, which followed protocols² developed by EPA's Industrial Environmental Research Laboratory at Research Triangle Park, North Carolina, characterized in detail the chemical composition of the three major streams discharged from the CAFB: boiler flue gas, spent lime sorbent, and ash from the sulfur recovery system.

The major environmental conclusions of the demonstration programs are summarized below:

- Nitrogen oxide emissions from the combustion of product gas generated by gasification of oil, lignite or bituminous coal were consistently below 100 ng/J (0.233 lb/10⁶ Btu) and were lower than NO_x emissions from natural gas combustion.
- During oil gasification the sulfur removal efficiency (SRE) varied between 60-90 percent. The SRE decreased markedly as the gasifier bed temperature was raised above 870°C (1600°F). Recycling of RESOXTM tail gas to the gasifier decreased SRE, probably as a consequence of steam and elemental sulfur entrained in the tail gas.

- Particulate emissions during oil gasification averaged 31.3 ng/J (0.073 lb/10⁶ Btu) based on the results of five separate EPA Method 5 tests. Opacity was generally below 20 percent. However, the higher than anticipated elutriation rate of bed material from the gasifier resulted in excessive particulate-build up in the boiler. Periodically this material would be blown off the boiler tubes by increasing the gas flow to the boiler which resulted in momentary jumps in the opacity.
- Particulate collected from the boiler flue gas showed neither appreciable toxicity nor any positive mutagenicity.
- Some mutagenicity was indicated by an Ames test on extracts of XAD porous polymer resin used to collect samples of gaseous organic compounds from the flue gas.
- Leachates collected from the spent stone and RESOXTM ash contained low metal and anion concentrations. Anion concentrations in both leachates were well below EPA Primary and Secondary Drinking Water Regulations.

REFERENCES

- 1a. Craig, J.W.T., et al. Study of Chemically Active Fluid-Bed Gasifier for Reduction of Sulphur Oxide Emissions. Final Report, June 1970 to March 1972. Prepared by Esso Research Centre, Abingdon, Berkshire, England, for the U.S. Environmental Protection Agency. EPA-R2-72-020. June 1972.
- b. Craig, J.W.T., et al. Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil-Second Phase. Prepared by Esso Research Centre, Abingdon, Oxfordshire, England, for the U.S. Environmental Protection Agency. EPA-650/2-74-109. November 1974.
- c. Craig, J.W.T., et al. Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil-Third Phase. Prepared by Esso Research Centre, Abingdon, Oxfordshire, England, for the U.S. Environmental Protection Agency. EPA-600/2-76-248. September 1976.
- d. Craig, J.W.T., et al. Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil-Fourth Phase. Prepared by Esso Research Centre, Abingdon, Oxfordshire, England, for the U.S. Environmental Protection Agency. EPA-600/7-79-066. February 1979.
2. Lentzen, D.E., D.E. Wagoner, E.D. Estes, and W.F. Gutknecht. IERL-RTP Procedures Manual: Level 1 Environmental Assessment (Second Edition). Prepared for the U.S. Environmental Protection Agency by Research Triangle Institute, Research Triangle Park, North Carolina. EPA-600/7-78-201, October 1978.

SECTION 2

INTRODUCTION

The Chemically Active Fluid Bed (CAFB) process was developed in the late 1960's by the Esso Research Centre, Abingdon (ERCA), England as a means to generate electrical energy from high sulfur, high metal heavy fuel oil.¹ During the early-and mid-1970's, the process was demonstrated at a 2.9 MW pilot plant to provide energy efficiently from both liquid and solid fuels while simultaneously limiting emissions of sulfur oxides, nitrogen oxides, vanadium, and nickel. Based on the successful pilot plant tests in England, the U.S. Environmental Protection Agency (EPA) and the Texas-based Central Power and Light Company (CP&L) commissioned Foster-Wheeler Energy Corporation (FWEC) to construct a 17 MW demonstration plant in San Benito, Texas. The agreement with CP&L, signed in May 1975, provided for equipment procurements, construction, and operation of the demonstration plant. The contract with EPA (Contract No. 68-02-2106) covered engineering design, labor for start-up, fuels and feedstocks, and environmental monitoring. Design of the unit, which was based on the ERCA design and operating experience, was initiated in January 1975, and construction was completed in January 1979. Over 130 contractors and subcontractors were employed during construction. A test program using heavy oil, lignite, and bituminous coal was conducted between October 1979 and June 1981. The contract with EPA ended in early March 1981 and covered only the portion of the test program using oil as the fuel. Lignite and coal tests (except for purchase of lignite and coal under the EPA contract) were funded solely by CP&L and FWEC.

In the CAFB process, oil or coal is fed continuously into a fluidized bed of lime maintained at 870°C (1600°F). Air is preheated and fed to the gasifier in substoichiometric proportions. The fuel entering the gasifier is vaporized, oxidized, and cracked to produce a low-Btu, low-sulfur

gas which, after passing through cyclones to remove particulate, is burned in a conventional gas boiler. Sulfur contained in the oil initially forms various gaseous compounds, which then react with the bed lime to yield solid calcium sulfide. Some of the fuel-bound trace elements also are captured by the bed stone. The sulfided lime is cycled to a regeneration unit where it is oxidized to produce calcium oxide, which is returned to the gasifier, and sulfur dioxide which is sent to a sulfur recovery unit. At the San Benito plant FWEC employed its proprietary sulfur recovery system-RESOXTM.

This report describes the operation of the San Benito demonstration plant during oil, lignite, and coal gasification. Section 3 describes the process and identifies key design parameters. Details of the design are available from the EPA Project Officer for this report. Actual operation is summarized in Section 4 and discussed in depth in Appendices A through C. Multimedia discharges from the demonstration plant were monitored by FWEC and GCA/Technology Division. The results of environmental monitoring and assessments are described in Section 5 and in Appendix D.

REFERENCES

- 1a. Craig, J.W.T., et al. Study of Chemically Active Fluid-Bed Gasifier for Reduction of Sulphur Oxide Emissions. Final Report, June 1970 to March 1972. Prepared by Esso Research Centre, Abingdon, Berkshire, England, for the U.S. Environmental Protection Agency. EPA-R2-72-020. June 1972.
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- d. Craig, J.W.T., et al. Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil-Fourth Phase. Prepared by Esso Research Centre, Abingdon, Oxfordshire, England, for the U.S. Environmental Protection Agency. EPA-600/7-79-066. February 1979.

SECTION 3

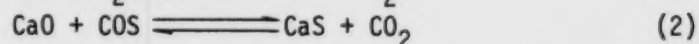
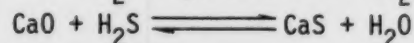
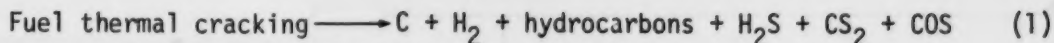
PROCESS DESCRIPTION

OVERVIEW

In the CAFB process, oil or coal is consecutively vaporized, oxidized, cracked, and reduced in a fluidized bed of lime to produce a low-Btu gas. This gas, from which a large portion of the sulfur has been removed by the lime, travels from the gasifier through cyclones for particulate removal and then into a conventional boiler for combustion. Under normal operating conditions, lime is continuously cycled between the gasifier and the regenerator where lime sulfided in the gasifier is oxidized to calcium oxide. Sulfur dioxide produced in the regenerator is sent to a recovery unit where sulfur is reduced to its elemental form. Spent lime is withdrawn from the regenerator for disposal on an intermittent or continuous basis. To maintain sulfur removal efficiency, an equivalent amount of limestone is added to the gasifier.

General Description and Chemistry of Gasification

The basic components of the CAFB process are the gasifier and regenerator. Figure 3-1 schematically illustrates the materials fed to and withdrawn from these unit operations. The quantity of limestone added to the gasifier is determined by the calcium/sulfur molar feed ratio necessary to provide the desired level of sulfur dioxide control. Substoichiometric air (~22 percent) is fed into the gasifier in order to partially oxidize the fuel oil and to produce a temperature of 871°C (1600°F) suitable for vaporization and cracking of the fuel. Flue gas from the boiler at approximately 171°C (340°F) is recirculated to the gasifier for temperature control. A low-Btu product gas is circulated from the gasifier to the boiler. The predominant reactions taking place in the gasifier are as follows:



The equilibria for these reactions are well to the right. Approximately five percent of the input limestone as calcium oxide is reduced to calcium sulfide on each pass of stone through the gasifier.

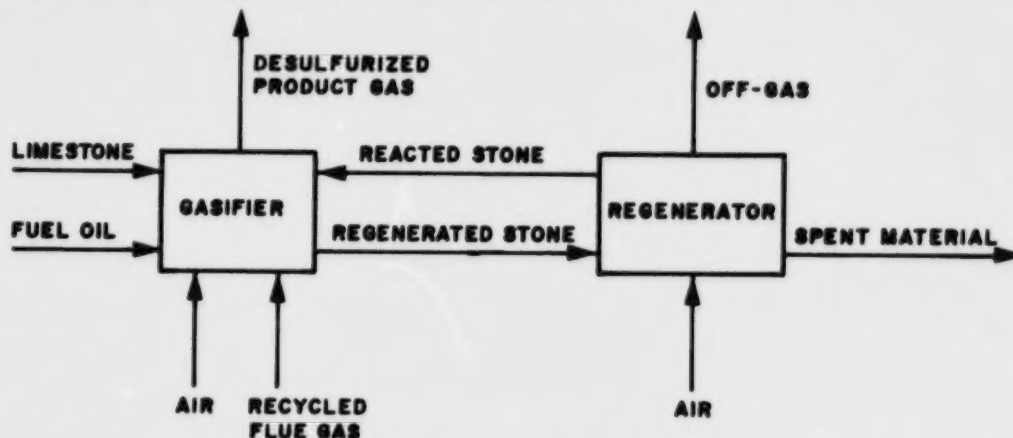
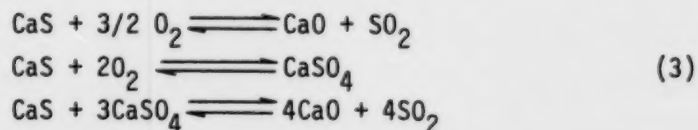


Figure 3-1. Generalized schematic of the CAFB.

General Description and Chemistry of Regeneration

Bed material composed of calcium oxide, calcium sulfide, calcium sulfate, carbon, and ash is fed to the regenerator from the gasifier where it reacts with a stoichiometric quantity of air by the reactions:



Carbon deposited on the stone during gasification is oxidized to carbon dioxide. These reactions maintain the regenerator temperature at the required 1040°C (1900°F).

The off-gas from the regenerator contains sulfur dioxide, carbon dioxide, and nitrogen derived from the influent air. Spent solid material consists of calcium oxide, calcium sulfate, calcium sulfide, and ash.

Off-gas is sent to a RESOXTM system for recovery of elemental sulfur, and the bulk of the spent solids are returned to the gasifier. Solids captured by the regenerator cyclone are conveyed to a solids cooler and storage bin.

Design Considerations

The design of the San Benito demonstration plant is based on the operating parameters developed by Esso Research Center Abingdon (ERCA). The unit was designed as a 17MW (based on oil) retrofit to an existing natural gas-fired boiler. Table 3-1 summarizes the design operating conditions for the No. 6 fuel oil defined in Table 3-2. The design also included provisions for gasification of coal.

Details of the design are presented in a Preliminary Process Design Manual prepared by Foster Wheeler Energy Corporation (FWEC) under EPA Contract No. 68-02-2106.² In lieu of isometric drawings and blueprints of the CAFB demonstration unit a detailed plastic model was constructed. The model is currently in San Benito, Texas. Pictures of the model can be obtained from the EPA Project Officer.

PROCESS DESCRIPTION²

The following process description is based on the design of the San Benito demonstration unit. A flow diagram of the CAFB process is presented in Figure 3-2.³ Each of the three distinct processes comprising the CAFB - gasification, regeneration, and sulfur recovery - are discussed separately below.

Gasification

The gasifier and regenerator are built into the same piece of equipment as shown in Figure 3-3.⁴ The gasifier/regenerator vessel is 3.66 m (12 ft) high and constructed of 7.94 cm (3.1 in) thick carbon steel with commercially available castable refractory insulation. The gasification side is 5 m by 3 m (16.6 ft by 9.9 ft) and is partially divided by a 23 cm (9 in) thick wall which forces the flow of the bed material from the regenerator outlet to the far wall of the gasifier and back to the regenerator inlet.

TABLE 3-1. DESIGN CONDITIONS

	Gasifier	Regenerator
Temperature, °C (°F)	871 (1,600)	1,040 (1,900)
Pressure, kPa-g (psig)	10.34 (1.5)	10.34 (1.5)
Limestone:		
Feed rate, kg/s (lb/h)	0.122 (970)	---
Static bed height, cm (in)	76 (30)	76 (30)
Fluid bed height, cm (in)	91 (36)	91 (36)
Bed drain rate, kg/s (lb/h)	---	0.0769 (610)
No. 6 Oil:		
Flow rate, kg/s (lb/h)	1.465 (11,631)	---
Sulfur, %	2.67	---
HHV, MJ/kg (Btu/lb)	42.8 (18,423)	---
Coal:		
Flow rate, kg/s (lb/h)	2.407 (19,100)	---
Sulfur, %	2.11	---
HHV, MJ/kg (Btu/lb)	31.07 (13,371)	---
Air:		
Flow rate, kg/s (lb/h)	3.932 (31,206)	0.5602 (4,446)
Temperature, °C (°F)	27 (80)	27 (80)
Flue Gas Recycle:		
Flow rate, kg/s (lb/h)	2.033 (16,132)	---
Temperature, °C (°F)	171 (340)	
Superficial Gas Velocity, m/s (ft/s)	1.4 (4.5)	1.5 (5.0)
Product Gas Flow Rate, kg/s (lb/h)	7.497 (59,500)	---
Regenerator Off-gas Flow Rate, kg/s (lb/h)	---	0.5935 (4,710)

Design value for total heat delivered to burners = 210×10^6 Btu/h (61.5 MW).
 Actual heat delivered was approximately 307×10^6 Btu/h (90 MW).

TABLE 3-2. ANALYSIS OF NO. 6 FUEL OIL^a

Component	Percent by weight
Carbon	84.43
Hydrogen	10.58
Nitrogen	0.37
Sulfur	2.67
Moisture	0.2
Ash	0.07
Oxygen (by difference)	1.68
TOTAL	100.00
Heating value (HHV)	
MJ/kg	42.8
Btu/lb	18,423
Specific gravity	0.9765

^aThis analysis was used in the design of the CAFB. An analysis of the oil which was gasified can be found in Section 5.

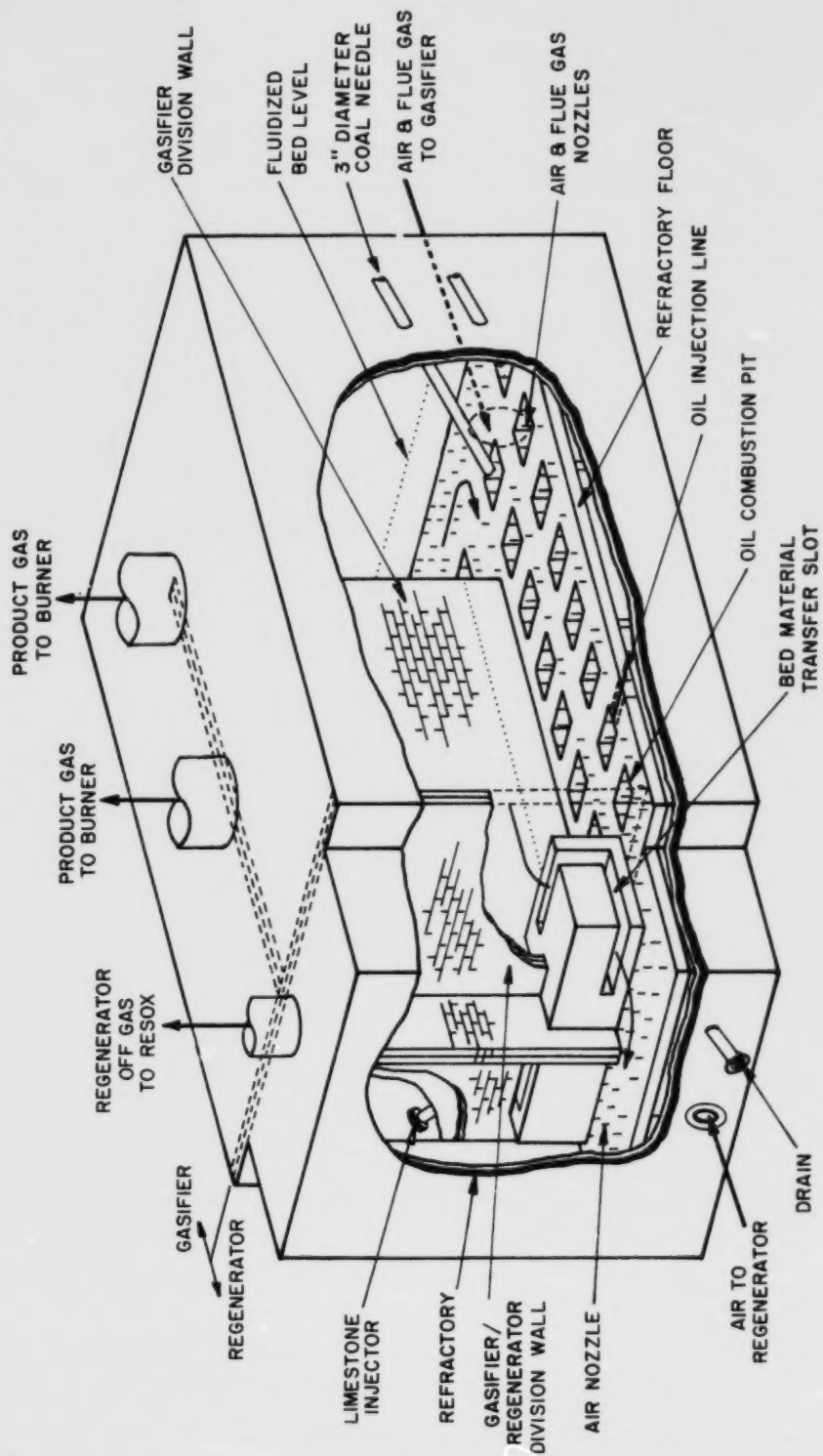


Figure 3-3. Gasifier/regenerator vessel.

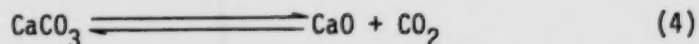
The gasifier is designed for gasification of coal or oil. It is equipped with 30 oil combustion pits and 24 coal feed pipes. The optimal temperature range for gasification is 850⁰-900⁰C (1560⁰-1650⁰F). At temperatures higher than 900⁰C (1650⁰F), the sulfur removal efficiency (SRE) decreases markedly. Air is supplied to the gasifier in substoichiometric proportion (approximately 22-23 percent) to limit both combustion and heat release. Flue gas is used as an inert, heat-absorbing medium to fluidize the bed while controlling the overall gasifier temperature. The velocity with which the air/flue gas passes through the bed and the depth of the bed affect gasifier performance. A high SRE is possible with a 1.22-1.83 m/s (4-6 ft/s) velocity range. A higher fluidizing velocity will blow an excessive amount of fines into the product gas cyclones. The optimal static bed depth should be greater than 38.1 cm (15 in) with an expanded height of 91.4 cm (36 in) to 96.5 cm (38 in).

The gasifier has a total of 590 air/flue gas nozzles. The air/flue gas nozzles act as a distributing system in the bottom of the gasifier bed for fluidizing the bed material. Equal distribution of these nozzles over the entire gasifier bed area is important for maximum fluidizing efficiency. The nozzles are of two sizes. The smaller ones are located in the oil combustion pits. The remaining 470 are molded into the floor of the gasifier.

During oil gasification, high-sulfur fuel oil is preheated to 121⁰C (250⁰F) and withdrawn from the oil storage tank. The oil is divided equally between the two oil distribution headers located at each side of the gasifier vessel. Each header supplies 15 oil injection nozzles which are located in the bottom of the 15 oil combustion pits. The combustion pits are 0.305 m (1 ft) square and 12.7 cm (5 in) deep. Each pit has four air/flue gas injection nozzles and one oil injection nozzle. As the oil burns, it supplies heat to vaporize and crack the remaining oil. The organic sulfur originally present in the oil forms a variety of sulfur compounds which react with the lime to form calcium sulfide, carbon dioxide, and water. In addition to the possible 90+ percent fuel-sulfur capture in the lime bed, tests indicate that 100 percent of any sulfur dioxide or hydrogen sulfide in the flue gas recycled to the gasifier section can be absorbed in the bed.

Efficient coal gasification is largely dependent on the coal particle size. Prior to gasification, the coal is transferred from the storage silo to a crusher. The coal is crushed to 100 percent through 1.3 cm (0.5 in), 88 percent through 6.35 mm (0.25 in), and 18-20 percent through 30 mesh screen. The crushed coal empties into a surge hopper equipped with dual outlets. Each outlet supplies a weigh feeder that feeds the coal through a rotary-air lock valve and into one of two pressurized distributing conveyors located on each side of the gasifier vessel. Each pressurized conveyor contains two vibrating tables which feed the coal through coal feed pipes arranged throughout the gasifier to ensure even distribution of the coal. There are twelve coal feed pipes each 7.7 cm (3 in) in diameter on each side of the gasifier. Each side has two rows of six needles. One row deposits the coal 0.99 m (3.25 ft) from the gasifier wall, and the second or inner row deposits the coal 0.30 m (1 ft) from the wall. All the coal feed pipes enter the gasifier at a 30-degree downward angle with the outlet of the needle approximately 7.6 cm (3 in) above the refractory floor. Good distribution and mixing of the coal with the hot limestone bed are necessary to attain efficient sulfur removal and gasification.

The fluidized lime bed is made up of regenerated lime and fresh limestone. The fresh limestone is transferred automatically from the fresh storage hopper to a surge bin located next to the gasifier. The bottom of the surge bin has a 1.2 m (4 ft) diameter bin activation device with a 15.2 cm (6 in) rotary air-lock valve attached. This valve can feed fresh limestone into the gasifier at approximately 0.12 kg/s (970 lbs/hr). The limestone feed rate is geared to maintain a calcium/sulfur molar feed ratio of 1:1 (calcium in the fresh stone to sulfur in the oil feed entering the gasifier). Fresh limestone is calcined in-situ according to the following reaction:



A small percentage of bed material escapes with the gasifier product gas. This bed material is captured by one of two cyclones located above the gasifier. Fines captured in the regenerator and product-gas cyclones

collect in the holding hopper and are removed through double-gate discharge valves. The captured fines are transported to the solids cooler by a refractory-lined pipe and a gas pulser. The cyclone fines are separated from the conveying line to the solids cooler by rotary valves whose function is to prevent excessive back flow of conveying gas up the diplegs which would reduce cyclone efficiency and cause the diplegs to overheat. The solids cooler takes the 1038°C (1900°F) bed material, composed of calcium oxide, calcium sulfate, calcium sulfide, and free carbon, and reduces the final temperature to 177°C (350°F). The cooler is made of a 6.35 mm (0.25 in) thick 316 stainless steel deck, a 316 stainless steel hood with removable inspection ports, an adjustable weir for variable material depth, and cooling air inlet and outlet spouts. Approximately 0.61 m³/s (1300 ft³/min) of air at 38°C (100°F) is required for cooling. The cooler outlet spouts feed a manifold which in turn feeds the dust-collecting equipment. The dust-collecting equipment captures any elutriated bed material. The cooled material discharges to a rotary valve which feeds into an air stream for pneumatic transport to the elevated spent-sorbent silo. Fresh limestone is added to the gasifier to replace the lost bed material and to maintain steady-state conditions within the entire system.

The gasifier and regenerator are separated by a wall which serves two major functions. The first is the separation of the reducing atmosphere in the gasifier from the oxidizing atmosphere in the regenerator. The second is the transfer of bed material between the gasifier and regenerator through two slots cast in the lower part of the wall, one on either side of the gasifier division wall. The transfer conduits are arranged so that they will generate a circulating flow from the gasifier to the regenerator and back. A set of flue gas nozzles is provided in each of the conduits to keep the material flowing through the transfer slots.

Initial test runs of the demonstration unit utilized flue gas to pulse bed material through the regenerator/gasifier transfer slots. The result was continual clogging of the slots. There were two reasons for this. Approximately 8 percent of the flue gas is water vapor.⁵ When the

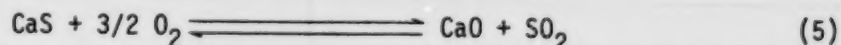
flue gas is routed to the gasifier/regenerator vessel, it cools, and the moisture present condenses. This condensation agglomerates the bed material in the slots. In addition, the oxygen present in the flue gas (approximately 7.5 percent of the flue gas is oxygen) reacts with the bed material to form a mixture of calcium sulfide, calcium sulfate, and calcium oxide. This mixture has a eutectic point below the operating temperature of the gasifier and regenerator. This increased the agglomeration of the bed material and also resulted in pluggage. To circumvent this problem, nitrogen was used in the later runs to pulse the bed material through the transfer slots.⁶

Product gas exits the gasifier chamber through two 76.2 cm (30 in) diameter product gas outlets located on top of the gasifier division wall. The flow rate of the product gas is approximately 7.5 kg/s (59,500 lb/hr). The 870°C (1600°F) essentially sulfur-free product gas is ducted through two cyclones before being burned in a conventional steam generator.

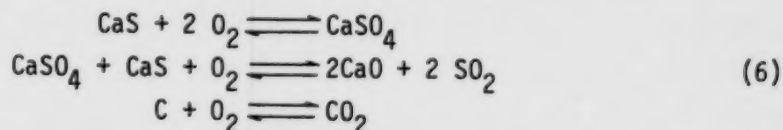
Regeneration

The sulfur removal efficiency of the CAFB is limited by a decreasing ability of the bed material to adsorb sulfur as the sulfur concentration increases. In order to maintain the SRE of the bed, sulfur-laden bed material must be removed from the gasification zone and replaced with sulfur-free lime. Sulfur-laden bed material is pulsed through the slots in the gasifier/regenerator division wall into the regeneration chamber. The regenerator vessel is 2.3 m by 1.1 m (7.5 ft by 3.5 ft) and has a plan area of 1.8 m² (19.3 ft²). Seventy-eight air nozzles are used to fluidize the bed material to an expanded height of 91.4 cm (36 in). The optimal superficial gas velocity in the regenerator is 1.4-1.6 m/s (4.6-5.3 ft/s). The optimal regenerator operating temperature is 1040° (1900°F) with an outside skin temperature of approximately 93°C (200°F).

The sulfur-laden gasifier bed material is reacted with air to regenerate calcium oxide (lime) from the calcium sulfide in the bed material. This reaction releases a concentrated sulfur dioxide off-gas. Other components of the regenerator off-gas are carbon dioxide and nitrogen. The primary lime regeneration reaction is:



Regenerator temperature is maintained by combustion of carbon in the material carried over from the gasifier and by the calcium sulfide combustion reactions, which are exothermic. These reactions are shown below.



Off-gas exits the regenerator through a 61 cm (24 in) conduit, located above the regenerator division wall, at a flow rate of about 0.6 kg/s (4,710 lb/hr). The off-gas goes through a cyclone designed for an inlet particulate loading of 16 g/m³ (7 gr/ft³) before entering the RESOXTM system. The regenerated lime then reenters the gasifier through a transfer conduit connecting the gasifier and the regenerator.

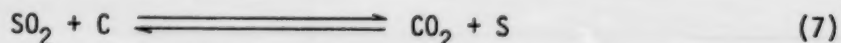
The original regenerator design included a drain pipe to remove a fraction of the bed material when needed. The removed bed material would be replaced with fresh limestone in the gasifier to maintain a calcium/sulfur ratio of 1:1. It was discovered during initial runs of the CAFB process that enough bed material was leaving the system with the gasifier product-gas and the regenerator off-gas to offset the fresh limestone feed and still maintain a calcium/sulfur ratio of 1:1.

Sulfur Recovery - The RESOXTM System

The RESOXTM system, a proprietary sulfur recovery process developed by FWEC, reduces sulfur dioxide contained in the regenerator exhaust gas to elemental sulfur which is then condensed to a liquid. Prior to entering the reactor vessel, the regenerator off-gas is stripped of its fines by a cyclone collector and cooled to approximately 650°C (1200°F) by steam or

water injection. Inside the RESOXTM reactor vessel the sulfur dioxide-rich gas passes through a slowly moving bed of anthracite coal, in which sulfur dioxide is reduced to gaseous sulfur and the coal is converted to ash.

The overall reduction reaction is:



The reactor vessel off-gas is passed through a counterflow wash for sulfur condensation. The stripped gas is then drawn through a tail-gas fan, which transports the gas to the flue-gas recirculation fan. The latter pressurizes and injects the RESOXTM tail-gas and the steam generator exhaust stack gases back into the gasifier for temperature control. Sulfur compounds remaining in the tail gas are absorbed by the bed.

Pollution Control Equipment

The San Benito demonstration plant uses particulate collection devices for both process control and air pollution control. Two cyclones intercept product gas from the gasifier before the gas enters the boiler. These cyclones serve two purposes: they remove particulate matter which could decrease boiler efficiency and plug product gas nozzles, and they reduce the amount of particulate matter released to the atmosphere in the boiler flue gas. Product gas enters the two parallel cyclones with a velocity of approximately 23 m/s (75 ft/s). The cyclones are designed for an inlet particle loading of 13.3 g/m³ (6 gr/ft³) with an efficiency of approximately 97 percent. The cyclones empty into a hopper from which material is sent to the spent stone cooler.

A cyclone is also used to capture particulate matter from the regenerator off-gas stream. This cyclone, which is designed for an inlet particulate loading of 16 g/m³ (7 gr/ft³) with an efficiency of 97 percent, is necessary to ensure proper operation of the RESOXTM sulfur recovery system. The collected fines are stored in a hopper under the cyclone and periodically sent to the spent stone cooler.

There are four baghouses and two filter receivers used to control fugitive particulate emissions from the San Benito plant. One baghouse is located over the fresh limestone storage tank. Fresh limestone is delivered by truck and unloaded into the storage tank via a pressurized

pneumatic transport system. Baghouse filters are located at the top of the hopper to control the transfer air vented to the atmosphere. This baghouse contains 14.7 m^2 (160 ft^2) of 16-oz Dacron filter bags which are cleaned by a reverse air flow. The fresh limestone is withdrawn from the fresh storage hopper and conveyed via a 7.6 cm (3 in) diameter pneumatic transfer line to a filter receiver consisting of 8.36 m^2 (90 ft^2) of 16-oz Dacron filter. The filter receiver empties into the feed hopper located adjacent to the gasifier.

Crushed anthracite coal is delivered to a receiving hopper for use in the RESOXTM reactor. The coal is transferred from the RESOXTM receiving hopper to the RESOXTM coal storage tank by way of a Redlar elevator. To control the dust created by this movement, a baghouse is located above the RESOXTM coal storage tank.

Another baghouse filter is located between the tail gas fan and the recycle gas blower. This baghouse removes particulate matter suspended in the RESOXTM system tail gas. The tail gas is used as an inert gas stream to fluidize the bed material in the gasifier and regenerator and to control the gasifier and regenerator temperature.

Additional fugitive air pollution control equipment is incorporated into the spent lime handling operation. Cyclone fines are collected in the cyclones above the gasifier and the regenerator. These fines are transported to the solids cooler where their temperature is reduced by air flow to 177°C (350°F). After cooling the fines the air is passed through a baghouse filter prior to being vented to the stack. The cooled fines are conveyed pneumatically to a spent-bed-material storage silo. At the top of the storage silo is a vent filter made of 8.36 m^2 (90 ft^2) of 16-oz Dacron filter bags.

Combustion of Product Gas

Unit No. 4 of the La Palma Power Station located in San Benito, Texas was selected for the CAFB retrofit. The unit was designed to fire both natural gas and oil. It has a maximum rated steam capacity of 31.5 kg/s ($250,000 \text{ lbs/hr}$), delivering superheated steam at 476°C (835°F) and 4.65 MPa-g (675 psig). Unit No. 4 is a 20 MW generator

which delivers 82 percent of the fuel heating value as steam when firing natural gas. It is equipped with four ring burners for natural gas firing and four steam atomizing oil burners for oil combustion. For the CAFB retrofit two larger capacity burners were installed, replacing the two existing bottom burners, to accomodate the CAFB product gas. Natural gas guns were installed inside these two product gas burners, enabling the steam generator to operate at half load while firing only product gas or at full load while firing both product gas and natural gas or natural gas only.

The EPA contract with FWEC called for the design of a CAFB unit which would deliver a product gas capable of generating 10 MW of electricity. The CAFB was overdesigned to provide 17 MW of generating capacity but achieved 22 MW during oil gasification.⁷ The maximum load for coal gasification was never attained because of problems with the coal feed system.

REFERENCES

1. Craig, J.W.T., G.L. Johnes, Z. Kowszun, G. Moss, J.H. Taylor, and D.E. Tisdall. Chemically Active Fluid-Bed Process for Sulphur Removal During Gasification of Heavy Fuel Oil-Second Phase. Esso Research Centre, Abingdon, Berkshire, England. U.S. Environmental Protection Agency, Research Triangle Park, North Carolina. Report Number EPA-650/2-74-109, November 1974.
2. Chemically Active Fluid Bed Process (CAFB) Preliminary Process Design Manual. Prepared by Foster Wheeler Energy Corporation Livingston, N.J. for U.S. Environmental Protection Agency, Research Triangle Park, North Carolina. Contract Number 68-02-2106. April 1, 1976.
3. Werner, A.S., C.W. Young, W. Piispanen, and B.M. Myatt. Preliminary Environmental Assessment of the Lignite-Fired CAFB. Prepared by GCA/Technology Division, Bedford, Massachusetts for U.S. Environmental Protection Agency, Research Triangle Park, North Carolina. EPA-600/7-79-048, February 1979.
4. McMillan, R.E. and F.D. Zoldak, "A discussion of the Chemically Active Fluid Bed Process (CAFB)". Oklahoma State University Frontiers of Power Technology Conference, Stillwater, Oklahoma, October 26 & 27, 1977.
5. Foster Wheeler Energy Corporation Monthly Technical Progress Narrative No. 62, for EPA Contract No. 68-02-2106, September 15, 1980. pg. 6-13.
6. Telecon. R.E. Sommer, GCA/Technology Division, (Chapel Hill, North Carolina), to Rich McMillan, Foster Wheeler Energy Corporation, (Livingston, New Jersey). February 14, 1983. Problems with the bed material transfer slots in gasifier/regenerator.
7. Foster Wheeler Energy Corporation Monthly Technical Progress Narrative No. 56; for EPA Contract No. 68-02-2106, March 17, 1980.

SECTION 4 PLANT OPERATION

INTRODUCTION

The purpose of this narrative is to describe the test runs of the CAFB process units which were retrofitted to the La Palma power station. It begins at the conclusion of the construction and calibration effort as the various feedstocks arrived on site and actual test work started. The narrative conveniently breaks down into three parts:

1. Combustion Tests,
2. Heavy Fuel Oil Gasification Tests, and
3. Oil/Lignite/Coal Gasification Tests.

Of necessity, the first part is by far the shortest since it only covers one test. The other two parts overlap to a certain extent because no test was exclusively on one fuel.

COMBUSTION TESTS

The CAFB gasifier plant, retrofitted to the No. 5 boiler (the No. 5 boiler supplies power to the No. 4 turbine/generator set) at the La Palma power station in San Benito, Texas, was completed by Christmas 1978, and equipment check out, calibrations, and preliminary trials started early in 1979. By this time most of the feedstocks, i.e. No. 6 fuel oil, lignite, and limestone for the gasifier had been procured and, in some cases, had been delivered to the site or local storage. These feedstocks are described in the first part of Appendix A. By March 1979, the preparatory checks and calibrations had been carried out, the refractories had been dried and fired, and preliminary cold fluidization trials had been completed. The cold fluidization trials described in Appendix A showed that fluidization in both the gasifier and the regenerator was satisfactory,

but the initial lift-off could be a problem. This confirmed the small scale results obtained at the Esso Research Center-Abingdon (ERCA) and by Foster-Wheeler Energy Corporation (FWEC) in laboratory tests.

Combustion trials were initiated in mid-May 1979 using a start-up procedure tried out successfully by ERCA and also by FWEC at a different location. Several attempts were made and numerous problems overcome before successful ignition and oil combustion were established for a short time in one half of the gasifier bed. Details of the combustion trials are given in Section 3 of Appendix A. Some equipment repairs and modifications (e.g. fitting shock absorbing mountings to reduce the vibrations caused by the operation of the coal feed table) were carried out before a second oil combustion attempt was made in early June. This attempt was successful, and steady combustion of No. 6 fuel oil was established. Owing to the shallow bed used (25 cm (10 in)), considerable overbed burning ensued, causing very high overhead gas temperatures and leading to the failure of the metal casing on one of the product gas cyclones. Repairs to the cyclone caused a considerable delay, and the next combustion trial could not be carried out until October 1979. Then after two abortive attempts, steady combustion of No. 6 fuel oil was established in the gasifier and was maintained for 66 hours.

A change over to gasification was postponed to a later date because the cyclone skin temperatures were high. In order to prevent a failure, the plant was shut down to install an air cooling system in the region of the annulus of each product gas cyclone.

A summary of the main events during this stage of the program is given in Table 4-1.

TABLE 4-1. COMBUSTION TESTS: MAIN EVENTS

Time	Main events
December 1978	Mechanical construction complete
January 1979	Refractories installed
February 1979	Drying, firing in, calibration
March 1979	Cold fluidization
May 1979	Start-up attempts
June 1979	First combustion test, aborted
October 1979	First combustion test, successful

OIL GASIFICATION

A number of oil gasification runs were carried out between December 1979 and May 1980 and are described in detail in Appendix B. Below is a summary of the salient features of these test runs; the main events are listed in Table 4-2.

Run 1 - December 1979

Following the installation of the cyclone cooling system, the CAFB gasifier was started and operated for almost 100 hours between December 16, and December 21, 1979. The modes of operation were approximately as follows:

- 63 hours of combustion, first utilizing lignite to start up, followed by steady combustion of No. 6 fuel oil. During this period CAFB contributed about 15 percent of the gross heat input into the boiler.
- 32 hours of gasification of No. 6 fuel oil with no regeneration.

Despite the severe solids carry-over into the boiler, the thermal efficiency of the boiler during product gas firing remained approximately the same as the thermal efficiency during natural gas firing. During the first 24 hours of the run, No. 6 fuel oil was gasified at a rate of 3311 kg/hr (7300 lb/hr). The product gas maintained a steady 10 MW with only one igniter on natural gas (0.3 MW). At the end of the steady gasification period, the number of oil guns in service was gradually reduced from 28 to 11. It was not possible to reduce the number of oil guns below 11 without reducing the oil flow, since at that point the oil flow control valve was fully open. There was no reduction in the sulfur removal efficiency (SRE) until the number of oil guns in service was reduced below 18. At 11 guns ($1.3 \text{ m}^2/\text{gun}$ or $14 \text{ ft}^2/\text{gun}$), however, the reduction in SRE was about 5 percent. As the run progressed and sulfur built up on the gasifier bed, the SRE deteriorated somewhat, while the stack opacity readings improved, implying that adsorption of sulfur strengthens the calcined limestone and reduces its capacity for further sulfur adsorption.

TABLE 4-2. OIL GASIFICATION TESTS: MAIN EVENTS

Time	Start-up/combustion	Gasification	Regenerator	RESOX™
December 1979	No start-up problems, 63 hr. combustion.	Run 1: 32 hr. gasification 24 steady, 8 hr. oil gun tests. Sulfation	Not in use	Not in use
January 1980	Sulfated bed used with no problems. Short combustion only.	Run 2: 77.5 hr. gasification 2 at M.C.R. of 22 MW, rest at steady 10 or 15 MW. Problems with sulfation	Not in use	Not in use
April 1980	Start-up problems: coarse fresh stone	Run 3: 144 hr. gasification of which 22 hr. product gas only	Not in use	Not in use
July 1980	Start-up problems: coarse fresh stone, 5 days on combustion.	Run 4: 90 hr. gasification RESOX™ tail gas fan problems	90 hr.	17 hr.

Run 2 - January 1980

The gasifier was preheated on January 26 using the sulfated bed from Run 1, and by the evening of January 27, gasification started. Gasification of No. 6 oil was continued for approximately 77.5 hours during an 82.5 hour period starting at 8 pm on January 27, 1980. The gasification run was interrupted for 5 hours because of a boiler trip which occurred after 58.5 hours of gasification. Load varied mostly in the range 10-15 MW, but near the beginning of the run, a maximum continuous rate (MCR) of 22 MW was maintained for about 2 hours. Operations during gasification were smooth. Problems were encountered with transfer of bed material between the gasifier and the regenerator, and consequently, the regenerator was not used.

The reason for shutdown was failure of four of the six gasifier thermocouples. The shutdown was orderly with a controlled transfer of the load to natural gas, followed by sulfation of the gasifier bed. The sulfation was not very well controlled and confirmed ERCA's finding that during sulfation very careful control of gasifier oxygen is necessary to maintain the temperature below 925°C (1700°F) to prevent accidental regeneration.

Sulfur removal efficiency fell steadily (possibly because of excessive carbon deposition on the bed) during the 22 MW run, reaching just below 60 percent when the load was reduced. Over the next 10 hours, SRE recovered, reaching nearly 90 percent, and remained above 80 percent for the next 40 hours. There the bed sulfur was about 20 percent, and SRE decreased sharply. Stack opacity was below 20 percent during most of the run except during the 22 MW period and during some rapid load changes between hour 49 and 55 of gasification.

Boiler efficiency appeared to be slightly higher on CAFB product gas than on natural gas, despite a high flow of fines through the cyclones and consequent deposition on boiler pipes.

Run 3 - April 1980

The third gasification run was initiated just before midnight of April 7, 1980, and product gas combustion started at 17:20 the next day. This run continued for 144 hours in a span of 148 hours, interrupted by two gasifier trips both caused by inadvertant erroneous actions. During that time the boiler was fired with 100 percent product gas for 22 hours, while for the remaining 122 hours, up to 50 percent of the boiler heat was produced by natural gas. This ensured that in the event of a gasifier trip, the boiler could continue to be fueled. For a large proportion of that period the gasifier temperature and bed velocity were on automatic control.

The end of the run was unforced; gasification was discontinued at 22:18 on April 14 and was followed by sulfation and a short period of combustion.

This test run was essentially trouble free and was very similar to the previous runs, the only difference was glazing of bed material during a temperature excursion during the second gasifier trip. Since the bed was essentially inactive after this temperature excursion, the small amount of desulfurization must have been carried out by the newly added bed material, and there were some signs of recovery. The regenerator was not used since there were problems with bed transfer.

Run 4 - July 1980

The fourth gasification run was begun at approximately 12:27 p.m. July 29, 1980. The total duration of gasification was 90.4 hours out of a total time of 97.7 hours. For most of the run (slightly more than 72 hours), the boiler was dual-fired with product gas derived from oil and with natural gas, while for the remainder of the run (slightly more than 11 hours), only oil derived product gas was used. After 68 hours of gasification, the unit tripped and was down for a little over 7 hours. A hot gasifier restart was then initiated.

Transfer of material between the gasifier and the regenerator commenced when the gasifier went into a combustion mode 6 days prior to gasification. The regenerator temperature increased when gasifier material was introduced;

this increase was maintained by the introduction of oil. The operation continued until gasification commenced, at which time oil injection into the regenerator ceased. Temperatures were then maintained by combustion of carbon and sulfur on incoming bed material. The regeneration of incoming bed material probably started shortly after gasification commenced; however, because of problems with analytical equipment, the actual time was unknown. Wet chemical analyses of regenerator off-gas yielded 2 volume percent sulfur dioxide within a few hours of initial gasification. Nitrogen was used for bed material transfer between the gasifier and regenerator; no difficulty was encountered with material transfer throughout the run.

Regenerator off-gas was first introduced into RESOXTM in the 51st hour of gasification. RESOXTM continued in operation for approximately 17 hours, at which time high sulfur temperatures caused the sulfur pumps to trip on high amps (see Appendix B). This in turn caused RESOXTM and ultimately the regenerator, gasifier, and boiler to trip. Flow of regenerator off-gas to RESOXTM was begun again during the 92nd hour of operation. RESOXTM operation was terminated less than 1 hour later, because of problems with the RESOXTM tail gas fan.

OIL, LIGNITE, AND COAL GASIFICATION

Run 5 - September 1980

This run started most inauspiciously; the first start-up attempt was aborted because of a fire in the spent fines bag house and the consequent delay while new bags were installed. However, once started, this was probably the most successful run to date. The actual start (from a cold bed of fresh lime to steady oil combustion) took less than one day. Following this trouble-free start-up, the regenerator was started and was maintained hot mostly by stone circulation. The change over to oil gasification lasted about one minute, and a few minutes later the regenerator began operating. The RESOXTM system was put into service and, despite a faulty tail gas fan, operated successfully for most of the test run. Finally, combined lignite and oil gasification was carried out for a period of 2.5 hours during which 2.3 Mg (2.5 tons) of lignite were gasified. Gasifier,

regenerator, and RESOXTM performed well; however, RESOXTM tail gas injection into the gasifier appeared to upset the SRE. This effect was traced to a large steam content of the tail gas and possibly to the detrimental effects of entrained elemental sulfur on sulfur adsorption by the lime.

Run 6 - November 1980

Several unsuccessful attempts at start-up were traced to an underlying cause of poor limestone size distribution. With a new supply of limestone the start-up caused no problems. The change over to oil gasification was rapid and trouble free. After a period of oil gasification two combined runs of oil and lignite gasification were carried out. There were some problems with the coal injector blocking, but the run was terminated when a bearing on the fines cooler failed and it was discovered there was very little fresh limestone left. Stack opacity increased and stayed high until oil replaced lignite. Since these were exploratory runs, very little environmental data were collected.

Run 7 - Lignite - February 1981

Following the previous attempts at partial oil replacement with lignite (Runs 5 and 6), a gasification run of lignite only was attempted. This test yielded 3.4 hours of successful lignite gasification. During this time 13.6 Mg (15 tons) of lignite was gasified at an average rate of 4 Mg/hr (4.4 tons/hr). During the entire run a total of 22.3 Mg (24.5 tons) of lignite was gasified, but part of it was during a combined lignite/oil operation. Fuel utilization was about 900 kg/hr/MW (200 lb/hr/MW).

A number of new features were tried during this test run:

- a) Used sulfated gasifier bed material was retained from the previous run as the initial start-up charge. There were no start-up problems, and the time from the initial fluidization to steady oil combustion was about 2 hours, (i.e. as short as was reasonable without refractory damage).
- b) The gasifier bed was sulfated and isolated part way through the run to allow repairs to be carried out.
- c) The gasifier bed was sulfated at the end of the lignite gasification run without transferring back to oil gasification.

Boiler flames were well defined although somewhat hazy; the haze of very fine ash particles also caused high stack opacity readings. Desulfurization was not good during the run. This was probably the result of the high gasification temperature, but it also may have resulted from stripping of sulfur on the bed material, as the initial bed sulfur was high. In any case the conditions were changing throughout the run, and equilibrium conditions were not established.

There were no lignite injector blockages during this run; instead, damp lignite was hanging up in the lobes of the rotary feeders and in elevator buckets causing feeding and transfer problems.

Run 8 - Lignite - February 1981

This run was an attempt to repeat Run 7 but with the coal feed system operational. As far as its primary objective was concerned, the run was successful in that lignite alone was gasified for almost 2 hours, with regeneration, at a 5490 kg/hr (12100 lb/hr) lignite feed rate and no problems with the coal feed system. The lignite in the site silo had been air dried before use which accounted for the absence of sticking and bogging down. The lignite utilization rate was similar to that obtained in Run 7, about 770 kg/hr/MW (1700 lb/hr/MW), and the lignite/oil replacement ratio was about 2.3 lb/lb.

Problems occurred with combustion of carbonaceous fines in the cyclone drains and also in the spent solids cooler. The run was terminated prematurely because of overheating and possible ignition in the solids cooler. However, the shut down was orderly with a switch-over to oil gasification and sulfation of the bed.

Run 9 - Bituminous - March 1981

Initial bituminous coal gasification was achieved on March 11, 1981 at 21:12 and continued for a total of 3 hours in three operating periods of 1.3 hours, 0.8 hours, and 0.9 hours, (21:12 to 22:30, 22:40 to 23:26 and 23:25 to 00:32 on March 12). The two interruptions were caused by malfunctions of the 'C' and 'D' coal feed tables. Eventually both 'C' and 'D' coal feed tables were taken out of service, and since they were both

on the same (front) side of the gasifier, the unbalanced temperature distribution could not be tolerated, and the operation was switched back to oil gasification. A total of 15.9 Mg (17.5 tons) of coal was gasified during this test run, of which 8 Mg (9 tons) was during a coal-only operation.

Boiler flame was clear and bright, stack opacity averaged 20 percent, and nitrogen oxide emissions were low at 86.1 ng/J ($0.2 \text{ lb } 10^6 \text{ Btu}$). Carbon utilization was acceptable at 60 percent, but the SRE was less than 50 percent. Most of this coal gasification run was carried out at a gasifier temperature in excess of 1038°C (1900°F). This high temperature is comparable to the temperature at which regeneration occurs and thus probably caused the low desulfurization efficiency.

During the actual gasification run there were no operational problems, and the only plant or mechanical problems were those with the coal feed tables (coal backing up on the table or not feeding to the rotary feeders) and with product gas cyclone dipleg seals. The fines descending down the diplegs were hot and had a high carbon content, while the conveying gas leaking past the rotary seals contained sufficient oxygen to cause combustion of the carbon with consequent overheating of both the diplegs and the rotary seals.

The regenerator was not operational during this test run.

Run 10 - Bituminous - June 1981

During the afternoon and night of June 9 and 10, 1981, a second highly successful test run was carried out, during which bituminous coal was gasified for a total of 12.5 hours in two periods: 5.5 hours (15:36-21:15 on June 9) at a steady rate of 3900 kg/hr (8800 lb/hr) and 7 hours (22:40 on June 9 - 5:40 on June 10, 1981) at a slightly higher rate of 4220 kg/hr (9300 lb/hr). During this test run 50 Mg (55 tons) of bituminous coal was gasified; 46 Mg (51 tons) during coal only operation and the remaining 4 Mg (4 tons) during the oil to coal change-over and during a combined oil/coal gasification during a feed train upset. A feature item of this test run was a 100 percent coal gasification re-start after a plant trip.

The regenerator was not operational during this test run.

Run 11 - Bituminous - June 1981

Run 11, the last in this series, was carried out during the evening and night of June 22 and 23, 1981. West Virginia bituminous coal was gasified by itself for a little over 9 hours and in conjunction with oil, for an additional 3.5 hours. During that time 36 Mg (40 tons) of coal was gasified at a carbon conversion efficiency of 67 percent and a carbon utilization efficiency of 74 percent. The initial desulfurization efficiency was good, at about 70 percent, but fell off because the regenerator was not in use. The start-up gasifier bed was held over from the previous coal gasification run and had an appreciable sulfur load from the start.

CONCLUSIONS

Oil, lignite, and bituminous coal were gasified separately or together in a 17 MW CAFB installation on the No. 5 boiler at the La Palma power station in San Benito, Texas. The operating results confirmed the pilot plant results qualitatively, and in some cases quantitatively and confirmed the feasibility of retrofitting such a unit to the front end of an existing natural gas boiler without derating and without an appreciable loss in efficiency.

A number of areas in which improvements are necessary have been highlighted; in particular:

- Single stage cyclones for product gas clean up were at times inadequate and needed improvement.
- The coal feed system, as used, was very sensitive to coal physical properties and was prone to plugging up under some conditions.
- A burn-up cell for carbon on cyclone fines during coal gasification would be desirable.
- Some shortcomings were noticed with some plant items, e.g. rotary seals, valves, dampers, augers, and some filters when exposed to the somewhat severe and rugged conditions of the CAFB unit.

SECTION 5

ENVIRONMENTAL ASSESSMENT

INTRODUCTION

The CAFB process was evaluated to determine its capability to control flue gas criteria pollutant emissions and to estimate potential multi-media pollution impacts. Continuous monitors were used by FWEC to measure sulfur dioxide, nitrogen oxides, carbon monoxide, oxygen, carbon dioxide and opacity. In addition, periodic manual emission tests were conducted by FWEC for particulate, sulfur dioxide, and nitrogen oxides using EPA reference methods 5, 6, and 7, respectively. As a point of reference for emission values presented in this section, Table 5-1 lists the New Source Performance Standard (NSPS) for utility (<73MW) steam electric generators, promulgated in 1971 and in effect at the time construction of the demonstration plant was initiated.¹ The San Benito plant is smaller than the utility size cutoff and thus was not subject to the limitations in the table. Also, as an experimental system the plant received a waiver from the Texas Air Control Board from compliance with State air pollution regulations. Therefore, comparisons with NSPS emission limits presented in this section are illustrative only and are not to be construed as compliance evaluations.

Potential multi-media environmental impacts of the CAFB were the subject of a separate program conducted for EPA by GCA/Technology Division under subcontract to Catalytic, Inc. (EPA Contract No. 68-02-2695). This program characterized in detail the chemical composition of the three major discharge streams from the CAFB: boiler flue gas, spent lime sorbent, and ash from the sulfur recovery system.

Results from both the FWEC and GCA contracts are presented in this section. FWEC collected data throughout all 11 runs. The GCA measurement program was conducted during oil gasification Run 5 only. Detailed results on the GCA program are presented in Appendix D.

TABLE 5-1. NEW SOURCE PERFORMANCE STANDARDS FOR FOSSIL-FUEL-FIRED STEAM GENERATING UNITS CONSTRUCTED AFTER AUGUST 17, 1971

Fuel ^a	Nitrogen Oxides		Sulfur Dioxide		Particulate	
	ng/J heat input	lb/10 ⁶ Btu heat input	ng/J heat input	lb/10 ⁶ Btu heat input	ng/J heat input	lb/10 ⁶ Btu heat input
Oil	130	0.30	340	0.80	43	0.10
Lignite	260	0.60	520	1.20	43	0.10
Coal	300	0.70	520	1.20	43	0.10
Natural Gas	86	0.20			43	0.10

^aWhen fuel types are mixed, the standards for nitrogen oxides and sulfur dioxide are prorated.

Emissions data were collected at the boiler exit, at the induced draft fan exit, and at several locations on the stack, as are shown in Figure 5-1. Monitoring instrumentation used to collect data is listed in Table 5-2. Data from all three sampling locations are presented in this section when available. Although the stack emission values are used to judge compliance with NSPS limits, stack data were not always obtained. In addition, emission values measured in the stack were often considerably higher than the emission values determined at the boiler exit and the induced draft fan exit. This was typically the result of venting the regenerator off-gas directly to the stack whenever the regenerator was in service and the RESOXTM reactor was not. On those occasions boiler exit and induced-draft fan exit emission values are more representative of the process emissions than the numbers measured in the stack. The small differences between the emissions measured at the boiler exit and those at the induced draft fan exit are due to air leaking into the ducting which is under negative pressure. The differences between the two numbers generally are within the range of error of the data measured at either location.

A total of 11 test runs were conducted between November 1979 and June 1981. Three types of fuel were gasified during that time: high-sulfur high-metals residual fuel oil; west Texas lignite coal; and West Virginia bituminous coal. The first 4 runs focused on gasification of residual fuel oil. The fifth run was primarily an oil gasification run, however lignite was introduced as a fuel during the final three hours of gasification. The next 2 runs gasified lignite and the final 4 runs gasified bituminous coal. Coal and oil were fed to the gasifier at one time or another during all 11 runs. Prior to gasification coal would be combusted in the bed to raise the bed temperature to the desired level. In the first 5 runs oil gasification would follow combustion. Toward the end of the fifth run the first attempt to gasify coal was made. The remaining 6 runs were coal gasification runs. In a coal gasification run, gasification would be started with oil and then switched to coal. Whenever a problem with feeding the coal occurred, oil would be fed to the gasifier until the problem was resolved.

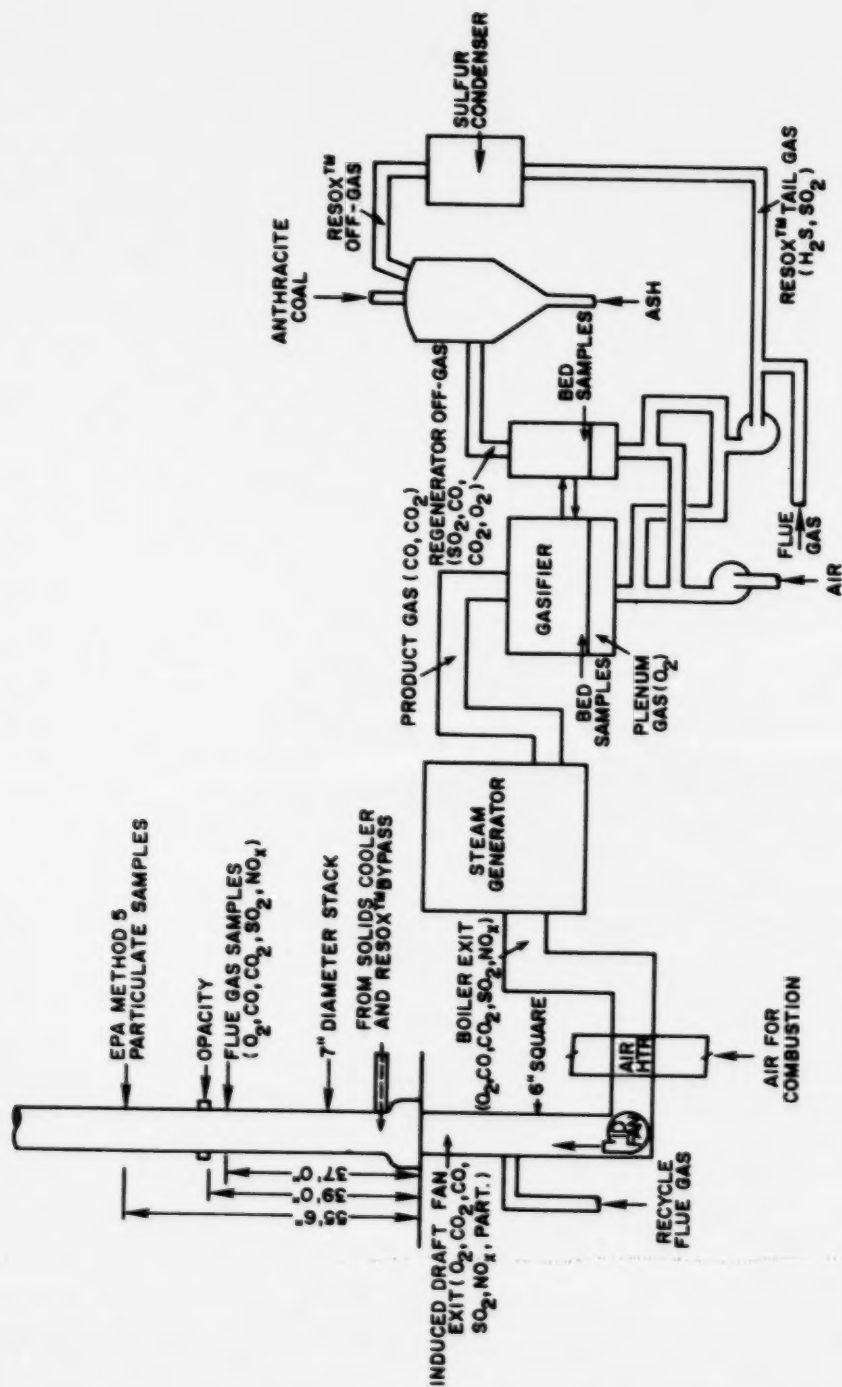


Figure 5-1. Sampling locations.

TABLE 5-2. INSTRUMENTATION FOR CONTINUOUS GAS STREAM ANALYSIS

Gas stream constituent	Instrument manufacturer	Principle of operation
Oxides of nitrogen	Thermo Electron Corporation	Chemiluminescence
Sulfur dioxide ^a	Lear Siegler	Ultra Violet Light Absorption -- Second Derivative Spectroscopy
Oxygen	Mine Safety Appliances	Nondispersive Infrared
Carbon monoxide	Mine Safety Appliances	Paramagnetic
Carbon dioxide	Mine Safety Appliances	Nondispersive Infrared
Opacity	Mine Safety Appliances	Nondispersive Infrared
	Lear Siegler	Light Scattering (double pass type)

^aThe Lear Siegler was used whenever it was operational. Both were used after the MSA was obtained in September 1981.

OIL GASIFICATION

An analysis of the No. 6 oil gasified during the first 5 test runs is listed in Table 5-3. The first gasification run in December 1979 lasted 34 hours. The second gasification run in January 1980 lasted 71 hours. During the second run a maximum continuous rating (MCR) test was performed for approximately 2 hours during which the boiler load was maintained between 21 and 22 MW solely on product gas.² The third gasification run in April 1980 lasted approximately 144 hours. During this run, several attempts were made to start the regenerator and RESOXTM reactor, neither of which was operated during the first two runs. When all efforts to start the regenerator failed, the run was terminated. The fourth gasification run in July 1980 lasted 91 hours. The regenerator was operational for 83 hours. The RESOXTM reactor was brought into service for a short time before the sulfur pump motors overloaded and tripped, causing the entire system to shut down. The fifth gasification run during September of 1980 lasted for 172.5 hours. The regenerator operated at the proper temperature for 130 hours, and the RESOXTM system was up for 115 hours.

Emissions data are presented below for the third, fourth, and fifth runs. The GCA test program was conducted during the fifth run. A summary of the results of that program are included under Run 5.

Run 3 - April 1980

Throughout the third gasification run, effluent gases were analyzed continuously for oxygen, carbon dioxide, carbon monoxide, sulfur dioxide, and nitrogen oxides from the boiler exit, the induced-draft fan exit, and the stack. Stack opacity was measured continuously, and two EPA Method 5 particulate analyses were performed. During this run a variety of gasifier and boiler operating conditions occurred. Data obtained during the run are grouped into 10 sets, each set corresponding to particular operating conditions. Averaged emissions data for each operating condition are presented in Table 5-4. On two occasions during the gasification run, gas samples were extracted from the induced-draft fan for analysis of nitrogen oxides, and on three occasions for analysis of sulfur dioxide. These

TABLE 5-3. FUEL OIL ANALYSIS

Components		Properties	
Ash	0.04%	Density	988 kg/m ³ (8.246 lbs/gal)
Carbon	86.07%	Gravity	11.4° API
Nitrogen	0.30%	Heating Value	42,254,300 KJ/m ³ (150,943 Btu/gal)
Hydrogen	10.20%	Kinematic Viscosity	3.372 stokes (156.3 SFS ^a @ 122°F)
Sulfur	2.43%		
Oxygen	0.90%		
Nickel ^b	9.5 ppm		
Vanadium ^b	39 ppm		
Chromium ^c	1.2 ppm		
Water (by distillation)	<0.10%		

^aSaybolt Furo1 second.

^bAnalysis performed by inductively coupled argon plasma (ICAP) and atomic adsorption (AA).

^cAnalysis performed by spark source mass spectrometry (SSMS).

analyses were performed using EPA Method 7 for nitrogen oxides and EPA Method 6 for sulfur dioxide. The results of these tests and the corresponding readings from the continuous analyzers are shown in Table 5-5. In all cases but one, the instrument reading was greater than the reference method, in one case by as much as 39 percent. However, in most cases, the continuous monitor was in rough agreement with the reference method (within 15 percent). In one case, that of the final sulfur dioxide analysis, the two determinations were within 2 percent.³

Except for test period 8, nitrogen oxide emissions are well below the NSPS for oil-fired boilers of 130 ng/J heat input ($0.3 \text{ lb}/10^6 \text{ Btu}$).¹ Sulfur dioxide emissions were generally below the NSPS of 340 ng/J heat input ($0.8 \text{ lb}/10^6 \text{ Btu}$).¹ A gasifier trip occurred during the fifth operating condition. Before the trip, the sulfur dioxide level had been increasing slightly, but after the trip, the sulfur dioxide level showed a significant increase. The as-read data for sulfur dioxide are misleading because of dilution of sulfur dioxide when natural gas is introduced as a fuel. Therefore, to give proper representation of the data, they were corrected for dilution. When the corrected data are observed, the increase in sulfur dioxide following the gasifier trip appears more dramatic.

Particulate emissions were significantly lower during the third run than in past runs, probably because of the decrease in fresh stone feed. The opacity was reasonably low throughout the run and remained below 20 percent from the ninth hour of gasification on, with several exceptions. These exceptions coincided with the efforts to remove particulate matter deposition on tubes by both increasing natural gas flow and, ultimately, flue gas flow over the tubes. The results of the two Method 5 analyses for total particulate indicated particulate emissions of 17.6 and 22.8 ng/J of heat input (0.041 and $0.053 \text{ lb}/10^6 \text{ Btu}$). These analyses were performed while product gas supplied 71 to 65 percent of total heat input to the boiler, respectively. Correcting these data for dilution resulting from natural gas combustion gives 24.9 and 34.8 ng/J of heat input (0.058 and $0.081 \text{ lb}/10^6 \text{ Btu}$) respectively, compared with the NSPS for particulate emissions of 43 ng/J of heat input ($0.10 \text{ lb}/10^6 \text{ Btu}$).¹

TABLE 5-5. COMPARISON OF WET CHEMICAL GASEOUS ANALYSIS DURING RUN 3 (PPM)
TO CONTINUOUS ANALYZERS

Test period	Gas analyzed	Concentration determined (ppm)	Average (ppm)	Standard deviation	Continuous analyzer reading (ppm)
1	Sulfur Dioxide	86.6 85.5 86.6	86.2	0.6	119
2	Oxides of Nitrogen	96.2 98.7 95.2 96.1	96.5	1.47	115
2	Sulfur Dioxide	254 254 252.5	253	0.87	285
9	Oxides of Nitrogen	167 175 172 180	173	5.2	151
9	Sulfur Dioxide	724 720 792	745	40.8	757

The initial Method 5 test, performed during the 81st and 82nd hours of operation (test period 6) resulted in an off-white (slightly yellowish) collection on the filter paper--very similar in nature to all past particulate test filters, indicating that the particulate emissions are primarily lime. However, the second Method 5 test, performed during the 135th and 136th hours of operation (test period 9) resulted in the collection of black carbon particulate on the filter, an indication of incomplete combustion.

Run 4 - July 1980

During the fourth run, samples of effluent gas from the boiler exit, induced-draft fan exit, and the stack were continuously drawn and analyzed for oxygen, sulfur dioxide, nitrogen oxides, carbon monoxide, and carbon dioxide. Averages of the data recorded during each of the 4 operating periods are shown in Table 5-6. A review of the data in this table reveals that a small amount of air leaked into the boiler exit samples. Oxygen values in the boiler exit samples were therefore higher than they should have been (higher than some of the O_2 values in the induced draft fan exit samples), and other gaseous concentrations were lower than they should have been (as a result of dilution). However, when mass flows of constituents are determined, the effects of dilution are eliminated.

As Table 5-6 demonstrates, nitrogen oxide emissions were lower than the NSPS (as they had been in previous runs), not exceeding 90.2 ng/J ($0.21 \text{ lb NO}_x/10^6 \text{ Btu}$). The lowest nitrogen oxide emissions, averaging 64.4 ng/J ($0.15 \text{ lb NO}_x/10^6 \text{ Btu}$), were found during the third operating period, in which the boiler was fired with product gas alone.

Sulfur dioxide emissions in most cases were well below the NSPS values. The sulfur removal efficiency (SRE) was approximately 80 percent (80 percent SRE corresponds to an emission of approximately 258 ng/J ($0.6 \text{ lb SO}_2/10^6 \text{ Btu}$)) during this run except during the third operating period when the RESOXTM tail gas was recycled to the bed. During the operations at San Benito, it was found that recycling the RESOXTM tail gas resulted in a significant decrease in SRE. Original testing at the pilot plant had indicated that RESOXTM tail gas could be recycled to the gasifier.

TABLE 5-6. AVERAGED EMISSIONS DATA FROM RUN 4 (OIL)

Test period	Sample location	Oxides of Nitrogen			Sulfur Dioxide			Carbon monoxide concentration (ppmv)	Carbon dioxide concentration (Vol %)	Oxygen concentration (Vol %)	Opacity (%)	
		Concentration (ppmv)	Standard deviation (ppmv)	ng/J	1b/10 ⁶ Btu	Concentration (ppmv)	Standard deviation (ppmv)					ng/J
1	BE ^a	129	7.6	90.4	0.21	149	24.4	146	0.34	42.02	7.9	6.82
	ID ^b	134	7.4	90.4	0.21	129	41.9	120	0.28	56.7	8.8	6.43
	ST ^c	105	10.1	86.1	0.20	436	109.6	490	1.14	300	8.2	8.77
2	BE	117	11.0	81.8	0.19	124	10.3	120	0.28	127	8.6	6.72
	ID	119	10.5	81.8	0.19	125	10.3	120	0.28	121	9.2	6.68
	ST	100	3.9	81.8	0.19	506	160	580	1.35	170	9.4	8.95
3	BE	100	11.0	73.2	0.17	320	57.5	327	0.76	176	11.4	7.01
	ID	92.2	29.8	64.5	0.15	316	54.5	314	0.73	179	10.8	6.66
	ST	82.1	5.1	68.8	0.16	338	189	404	0.94	188	11.2	9.13
4	BE	111	6.1	86.1	0.20	179	39.9	197	0.46	76.2	9.1	8.43
	ID	119	4.5	90.4	0.21	217	42.8	223	0.52	166	8.8	7.58
	ST	94.2	3.3	86.1	0.20	599	58.2	744	1.73	113	8.5	9.77

^aBoiler exit.^bInduced draft fan exit.^cStack.

The testing was conducted with gaseous hydrogen sulfide, reduced organic sulfides, and sulfur dioxide. It is believed that elemental sulfur in the tail gas was causing the problem. Future designs of the CAFB will probably not recycle the RESOXTM tail gas to the gasifier.

In normal integrated operation the regenerator off-gas would be ducted to the RESOXTM reactor for sulfur recovery. During operating periods 1, 2, and 4, the RESOXTM reactor was not in service, and the regenerator off-gas was vented directly to the stack. This can be observed by comparing the sulfur dioxide emissions from the boiler exit and the induced draft fan with the sulfur dioxide in the stack. During the third operating period, the regenerator off-gas was ducted to the RESOXTM reactor; thus, the difference between the sulfur dioxide from the boiler exit or the induced draft fan exit and the stack is much less pronounced.

Carbon monoxide emissions were low throughout the run. The highest concentrations of emissions were detected during the third operating period, when the boiler was fired with product gas only. This was not surprising, because the product gas flame is long and hazy, especially when compared with the natural gas flame, and fills virtually the entire furnace area, sometimes even extending into the convection passes. The carbon monoxide levels during this period, however, were relatively stable in the 170 to 190 ppm range, with an average of 176.5 ppm.

Stack opacity was measured continuously and recorded periodically. The opacity was consistently below 20 percent and was less than 10 percent for most of the run. Exceptions to these figures include the 34th hour of gasification, when opacity went to 100 percent, and the 37th hour, when it went to 80 percent. During these periods, the boiler firing rate was increased by the introduction of more natural gas in an effort to dislodge dust deposits from boiler tubes, superheater tubes, and air heater tubes. Whenever this procedure is enacted, large quantities of dust are dislodged and, in general, stack opacity is increased.⁴

OIL, LIGNITE, AND COAL GASIFICATION

Run 5 - September 1980

During the fifth gasification run, effluent gases were analyzed continuously for oxygen, carbon dioxide, carbon monoxide, sulfur dioxide, and nitrogen oxides from the boiler exit and the induced draft fan exit. Stack opacity was measured continuously, and three EPA Method 5 particulate analyses were performed. The averaged data from the boiler exit for the 11 test run periods of operation are presented in Table 5-7. Particular attention should be paid to the fifth and eighth sets. It was during these periods of operation that the boiler was operating solely on product gas.

GCA conducted its test program during the eighth test period, on June 30, 1980. This test program followed the procedures for environmental assessments developed by EPA's Industrial Environmental Research Laboratory. These procedures are described in Reference 5 and summarized in Appendix D. The quality assurance and quality control procedures specified in this manual were followed by GCA during the test program. The major focus of their test program was the stack gas. The flue gas was tested for inorganic, gaseous, and organic constituents. Two solid waste streams, the RESOXTM ash, and spent limestone, were also analyzed. In addition, leachates from the solid waste streams were analyzed. Bioassay tests were performed on the flue gas particulate and organic matter, the solid waste streams, and the solid waste leachates.

Inorganic Analytic Results --

Particulate matter and trace elements in flue gas

Samples of inorganic and organic pollutants in the CAFB flue gas were obtained using the Source Assessment Sampling System (SASS). The SASS train collects particulate matter and inorganic trace species in three cyclones with cut sizes of 10 μm , 3 μm , and 1 μm , a particle filter, and two impingers. In addition, an Andersen impactor was used to determine flue gas particle size.

Particle size data derived using a SASS train and Andersen impactor are presented in Figure 5-2. The particle size data suggest that the size distribution is bimodal, with one group of particles smaller than about 1 μm in aerodynamic diameter and a second group larger than 3 μm .

TABLE 5-7. AVERAGED EMISSIONS DATA FROM RUN 5 (OIL)

Test period	Sample location	Oxides of Nitrogen			Sulfur Dioxide			Carbon monoxide concentration (ppmv)	Carbon dioxide concentration (Vol %)	Oxygen concentration (Vol %)	Opacity (%)		
		Concentration (ppmv)	Standard deviation (ppmv)	ng/J	1b/10 ⁶ Btu	Concentration (ppmv)	Standard deviation (ppmv)					ng/J	1b/10 ⁶ Btu
1	BE ^a	163	15.0	96.0	0.223	34.9	25.7	28.4	0.066	67.5	9.1	4.3	6.5
2	BE	131	12.5	80.0	0.186	48.8	9.6	41.3	0.096	36.9	10.7	4.85	0
3	BE	111	19.7	68.8	0.160	81.7	10.9	70.6	0.164	62.5	--	5.30	2.7
4 ^b	BE	130	15.7	79.2	0.184	86.9	26.5	73.6	0.171	31.5	--	4.86	0.5
5	BE	107	4.9	68.8	0.160	550	114	489	1.14	42	8.8	5.05	13.1
6	BE	110	3.1	70.6	0.164	776	5.8	686	1.60	113	8.8	5.30	8.3
7	BE	143	4.9	84.3	0.196	603	84.1	492	1.14	56.7	--	4.1	14
8	BE	123	1.41	84.3	0.196	710	183	677	1.57	--	--	6.1	--
9	BE	116	5.3	80.0	0.186	690	101	657	1.53	--	--	5.8	14.9
10	BE	125	5.2	82.2	0.191	479	105	438	1.02	--	--	5.54	--
11	BE	129	26.2	91.2	0.212	562	118	550	1.28	459	--	6.14	60

^aBoiler exit.^bThe RESOXTM reactor went into service during the fourth test period. The top number is prior to the RESOXTM being in service. The bottom number corresponds to when the reactor was in service.

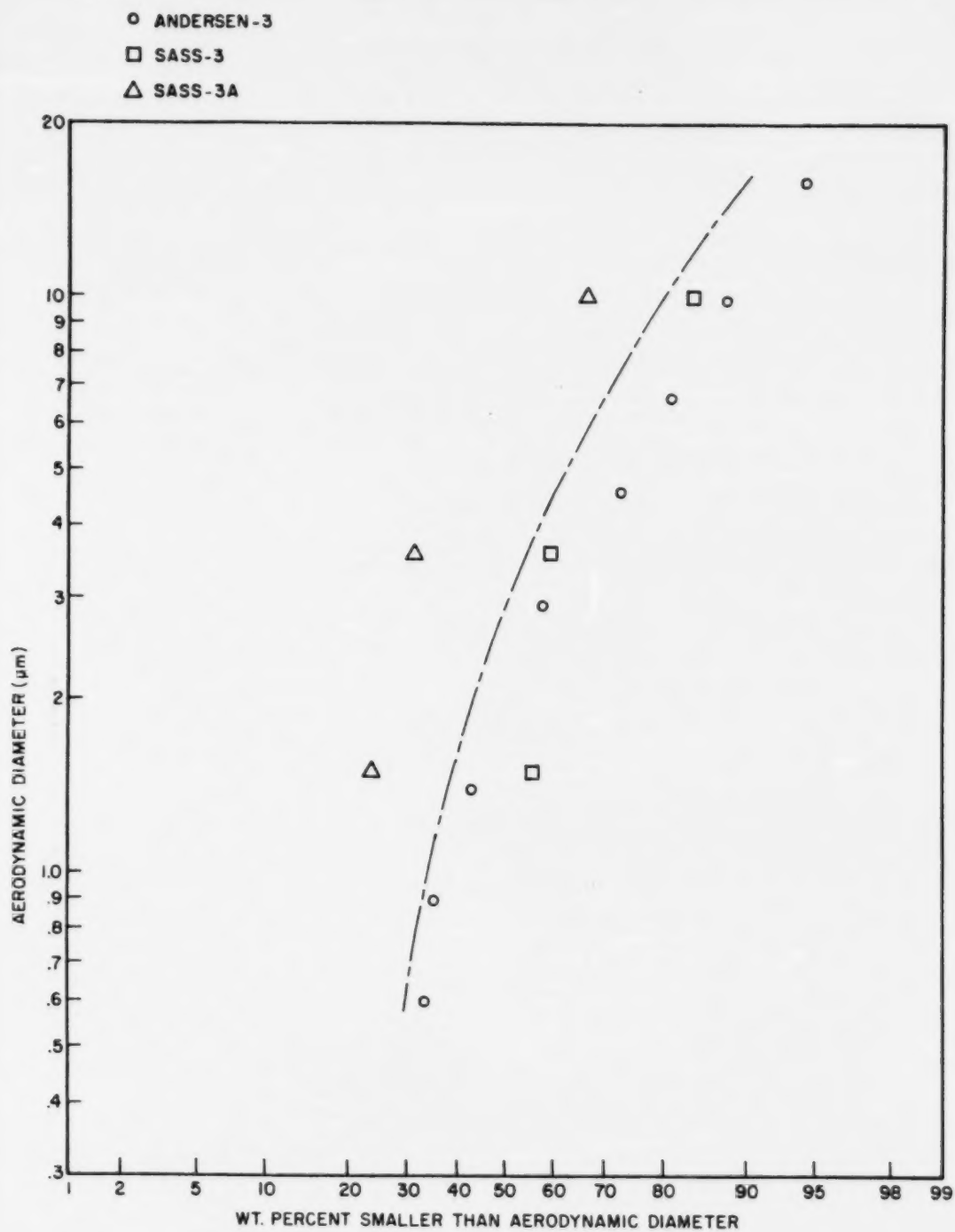


Figure 5-2. SASS train cyclone and Andersen impactor results.

Such a bimodal distribution was also observed in particulate emitted at the CAFB pilot plant.⁶ The bimodal distribution indicates that there are two processes by which particulate matter are generated in the CAFB. The larger particles are probably generated by the abrasion of the bed material and the smaller particles by the condensation of the ash present in the oil.

The particulate loading was measured during the eighth operating period with three separate EPA Method 5 tests. The results of that testing show a particulate loading of 54.1, 26.6, and 35.6 ng/J (0.126, 0.062, and 0.083 lb/10⁶ Btu). With the exception of the first test, these results are below the NSPS limits presented in Table 5-1. Because the boiler was operating on product gas alone during operating period 8, there is no correction for dilution.

The SASS train cyclone and filter samples were analyzed for trace elements using spark source mass spectrometry (SSMS). The SASS train impingers and the condensate from the SASS organic module were analyzed for trace elements using inductively coupled argon plasma spectrometry (ICAP) and atomic absorption (AA). The results of analysis for trace elements in flue gas samples can be found in Table 5-8. The reported results represent the concentrations measured in the total catch obtained using the SASS train (i.e., the cyclone catches, particle filter, and impingers). The individual concentrations measured in the cyclone catches, particle filters, and impingers can be found in Appendix D.

Solid waste streams

The bulk solid waste streams from the CAFB process were evaluated in two ways. Samples from the spent stone and RESOXTM ash were analyzed for trace elements by SSMS. In addition to this, distilled water leachates of the spent stone and RESOXTM ash were analyzed by ICAP, ion chromatography (IC), and AA (for Hg).

Results of the trace element analysis of the spent bed and RESOXTM ash are presented in Table 5-9. It is not clear if these two waste streams will be classified as hazardous under the provisions of the Resource Conservation and Recovery Act (RCRA). The waste streams are not listed

TABLE 5-8. TRACE ELEMENT CONCENTRATIONS IN FLUE GAS

Pollutant species	Discharge concentration ($\mu\text{g}/\text{m}^3$)	Pollutant species	Discharge concentration ($\mu\text{g}/\text{m}^3$)
Lithium	2 - 13	Rhodium	<1.0
Beryllium	<1.1	Palladium	<1.0
Boron	2 - 170	Cadmium	1.2 - 2.2
Fluorine	14	Tin	<1.0
Sodium	<5,600	Antimony	<3.0
Magnesium	43 - 330	Tellurium	<1.0
Aluminum	<930	Iodine	<1.0
Silicon	<14,000	Cesium	<1.0
Phosphorus	24	Barium	8.0 - 8.1
Sulfur	MC ^a	Lanthanum	6.0
Chlorine	8.0	Cerium	1.0
Potassium	MC	Praseodymium	<1.0
Calcium	MC	Neodymium	2.0
Scandium	<1.0	Samarium	<1.0
Titanium	7.3	Europium	<1.0
Vanadium	120	Gadolinium	<1.0
Chromium ^b	4	Terbium	<1.0
Manganese	23	Dysprosium	<1.0
Iron	270	Holmium	<1.0
Cobalt	7.3	Erbium	<1.0
Nickel ^c	42	Thulium	<1.0
Copper	29	Ytterbium	<1.0
Zinc	230	Lutetium	<1.0
Gallium	<1.0	Hafnium	<1.0
Germanium	<1.0	Tungsten	<1.0
Arsenic	6.5 - 9.4	Rhenium	<1.0
Selenium	25 - 28	Osmium	<1.0
Bromine	<1.0	Iridium	<1.0
Rubidium	1.0	Platinum	<1.0
Strontium	<89	Gold	<1.0
Yttrium	<1.0	Thallium	<1.0
Zirconium	1.0	Lead	9.5 - 11
Niobium	<1.0	Bismuth	<1.0
Molybdenum	<1.0	Thorium	<1.0
Ruthenium	<1.0	Uranium	<1.0

^aMajor component.

^bThis represents the chromium found on the 1 μm , 3 μm , 10 μm , and particle filter. The amount of chromium in the impingers can be found in Appendix D.

^cThis represents the nickel found on the 1 μm , 3 μm , 10 μm , and particle filter. The amount of nickel in the impingers can be found in Appendix D.

TABLE 5-9. TRACE ELEMENT CONCENTRATIONS IN THE SPENT BED AND RESOX™ ASH

Pollutant species	Discharge concentration (µg/g)		Pollutant species	Discharge concentration (µg/g)	
	Spent stone	RESOX™ ash		Spent stone	RESOX™ ash
Lithium	0.19	2.9	Rhodium	<0.31	NR ^a
Beryllium	0.089	<0.22	Palladium	<0.90	<0.18
Boron	14	<7.0	Cadmium	3.3	<2.0
Fluorine	37	19	Tin	<0.81	<0.70
Sodium	240	1,100	Antimony	<0.63	<0.24
Magnesium	6,700	190	Tellurium	<0.88	<0.18
Aluminum	300	41,000	Iodine	<0.39	<0.84
Silicon	23,000	51,000	Cesium	<0.065	0.28
Phosphorus	780	190	Barium	66	96
Sulfur	68,000	MC ^b	Lanthanum	32	4.2
Chlorine	150	<78	Cerium	22	3.8
Potassium	16	4,200	Praseodymium	2.2	<0.82
Calcium	530,000 ^c	830	Neodymium	8.2	<8.2
Scandium	9.8	1.3	Samarium	<1.4	<0.48
Titanium	130	590	Europium	<0.38	<0.14
Vanadium	2,800	15	Gadolinium	<0.71	<0.14
Chromium	8.6	5.9	Terbium	<0.61	<0.038
Manganese	2,300	13	Dysprosium	<0.72	<1.3
Iron	8,900	4,500	Holmium	<0.28	<0.082
Cobalt	6.9	4.0	Erbium	N	N
Nickel	260	7.3	Thulium	N	N
Copper	11	<18	Ytterbium	N	N
Zinc	120	<20	Lutetium	N	N
Gallium	10	5.7	Hafnium	<2.7	<0.40
Germanium	<0.41	<0.084	Tungsten	<1.1	<0.32
Arsenic	<0.26	1.0	Rhenium	N	N
Selenium	<0.61	<0.12	Osmium	N	N
Bromine	<1.4	<0.62	Iridium	N	N
Rubidium	<0.069	12	Platinum	<3.4	NR
Strontium	2,000	36	Gold	N	N
Yttrium	8.5	2.1	Thallium	<0.92	<0.17
Zirconium	<0.60	6.1	Lead	<1.6	<8.6
Niobrium	<0.34	<0.80	Bismuth	<0.75	<0.14
Molybdenum	<0.76	<1.1	Thorium	<2.1	<1.8
Ruthenium	<0.96	<0.22	Uranium	2.9	<0.62

^aNot reported.^bMajor component.^cThe concentration of calcium oxide in the spent stone was estimated by subtracting the measured concentrations of other major components from unity.

under subpart D; however, they could still be classified as hazardous if they are determined to be ignitable, corrosive, reactive, or EP toxic.⁷

Specific procedures are outlined in RCRA for determining if the waste material is hazardous under these classifications. The leaching procedure used in this environmental assessment differs from the leaching procedure used to determine EP toxicity. Therefore, the metal concentrations found in the RESOXTM ash and spent stone leachates cannot be used to determine if either of the solid waste streams are EP toxic. The results of the trace element analysis of the laboratory-generated leachates are shown in Table 5-10. The laboratory-generated leachates were also analyzed for various anions using IC. The results of this analysis are also presented in Table 5-10. Certain anion concentrations in drinking water are regulated under the EPA National Primary and Secondary Drinking Water Regulations.^{8,9} As a means of comparison these values are also presented in Table 5-10. The levels of the regulated anions are well below drinking water standards.

Organic Analytical Results --

Samples from the CAFB generally contained low levels of organic compounds. Organic analyses concentrated on the boiler flue gas stream. Limited analyses were performed on the solid waste streams, because the organic contents of the RESOXTM ash and spent stone were quite low. Therefore, in-depth analyses and interpretation of organic data were limited to the flue gas stream.

Flue gas

Organic compounds were captured in the porous polymer resin (XAD) and condensate modules of the SASS train and returned to the laboratory for analysis. Low resolution mass spectrometry (LRMS) was used to identify nonvolatile organic compounds present in the XAD extract. In addition the XAD resin was screened using high performance liquid chromatography (HPLC) with fluorescence detection. This procedure is especially useful for detection of polynuclear aromatic hydrocarbons (PNAs). Gas chromatography/mass spectrometry (GC/MS) was used to quantify specific PNAs. Figure 5-3 presents the GC/MS scan of the XAD extract. Table 5-11 presents the results of the HPLC and the GC/MS analyses.

TABLE 5-10. ANION AND TRACE ELEMENT CONCENTRATIONS IN SPENT BED AND RESOX™ ASH LEACHATE

Anion	Detection limit (ppb)	Sample analyzed		Maximum permissible concentration (ppb) ^a
		RESOX™ ash leachate (ppb)	Spent stone leachate (ppb)	
F ^{-b}	50	ND ^c	ND	1,400 - 2,400
Cl ^{-b}	100	400	ND	250,000
Br ^{-b}	500	ND	88,000	---
NO ₃ ^{-b}	500	ND	ND	10,000
SO ₄ ^{=b}	500	51,000	160,000	250,000
PO ₄ ^{=b}	500	ND	ND	---
Hg ^d	0.5	ND	ND	---
Sb ^e	15	ND	ND	---
As ^e	32	ND	ND	---
Be ^e	1.0	ND	ND	---
Cd ^e	1.0	ND	ND	---
Cr ^e	3.0	ND	ND	---
Cu ^e	2.0	ND	ND	---
Pb ^e	16	ND	ND	---
Ni ^e	5.0	ND	ND	---
Se ^e	22	ND	ND	---
Ag ^e	1.0	ND	ND	---
Zn ^e	1.0	24	3	---

^aFluoride and nitrate limits are set under EPA National Interim Primary Drinking Water Regulations (40CFR 141); chloride and sulfate under EPA National Secondary Drinking Water Regulations (40 CFR 143); -- indicates no maximum permissible concentration has been set.

^bConcentrations measured by ion chromatography.

^cNot detectable.

^dConcentration measured by atomic absorption.

^eConcentrations measured by inductively coupled argon plasma spectrometry.

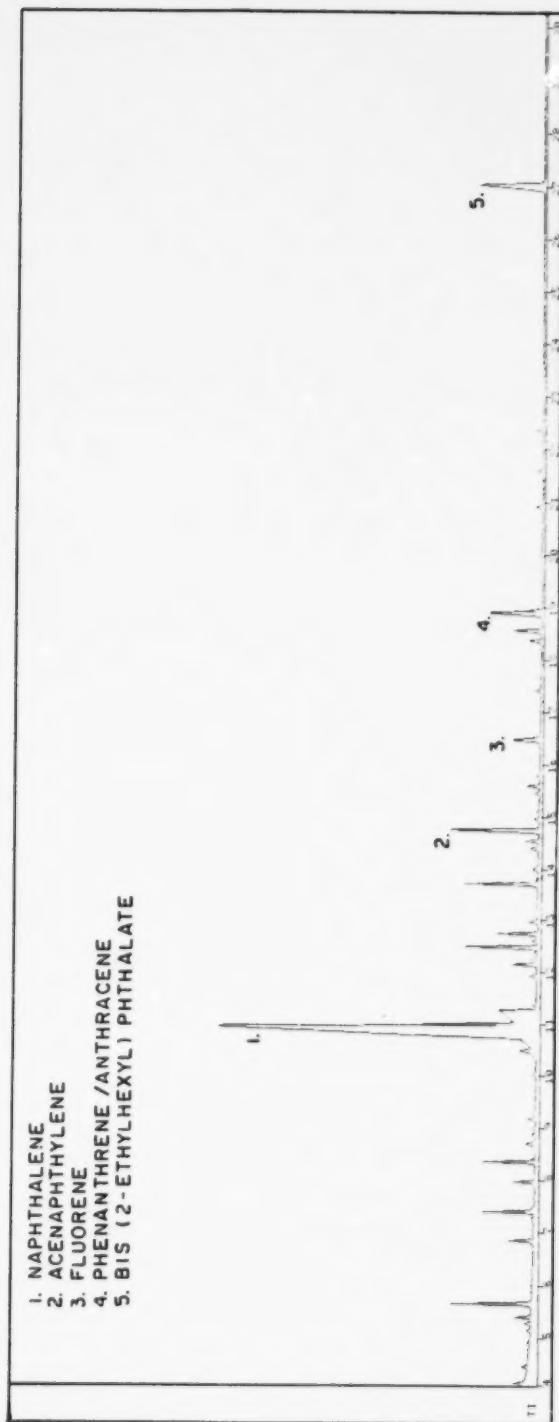


Figure 5-3. GC/MS scan of XAD adsorbent extract from Run 5.

TABLE 5-11. RESULTS OF HPLC/FLUORESCENCE AND GC/MS ANALYSIS FOR POLYNUCLEAR AROMATIC HYDROCARBONS IN CAFB FLUE GAS

PNA	Total SASS-train catch concentration ($\mu\text{g}/\text{m}^3$)	PNA	Total SASS-train catch concentration ($\mu\text{g}/\text{m}^3$)
Acenaphthene	<1	Dimethyl phthalate ^a	<1
Acenaphthylene	20	Diethyl phthalate	<1
Phenanthrene/anthracene ^{a,b}	60.4	Di-n-butyl phthalate	<1
Benz(a)anthracene ^a	<0.3	Dioctyl phthalate	<1
Chrysene ^a	<2.4	Butyl Benzyl phthalate	<1
Benzo(a)pyrene/benzo(e)pyrene ^{a,b}	<0.1	Bis(2-ethylhexyl)phthalate	61
Benzo(b)fluoranthene	<1	1,1'-Dichloromethyl ether	<1
Benzo(k)fluoranthene ^a	.011 - .012	2,2'-Dichlorodimethyl ether	<1
Benzo(g,h,i)perylene ^a	<0.2	2,2'-Dichlorodisopropyl ether	<1
Indeno(1,2,3-cd)pyrene ^a	<0.4	4-Bromophenyl phenyl ether	<1
Dibenz(a,h)anthracene	<9	4-Chlorophenyl phenyl ether	<1
Fluoranthene/pyrene ^{a,b}	.063 - .066	N-nitrosodimethyl amine	<1
Fluorene	9.4	N-nitrosodiphenyl amine	<1
Naphthalene ^a	99.	N-nitroso-d-N-propylamine	<1
2-Chloronaphthalene	<1	Hexa-chloroethane	<1
1,2-Dichlorobenzene	<1	Bis(2-chloroethoxy)methane	<1
1,3-Dichlorobenzene	<1	Isophorone	<1
1,4-Dichlorobenzene	<1	Hexa-chlorobutadiene	<1
1,2,4-Trichlorobenzene	<1	3,3'-Dichlorobenzidine	<1
Hexachlorobenzene	<1	Benzidine	<1
Nitrobenzene	<1	1,2-Diphenylhydrazine	<1
2,4-Dinitrotoluene	<1	Hexachlorocyclopentadiene	<1
2,6-Dinitrotoluene	<1	2,3,7,8-Tetrachlorodibenzo-p-dioxin	<1

^aHPLC/fluorescence was used; all other analyses were done with GC/MS.

^bThese species coelute under our experimental conditions.

Solid waste streams

The two solid waste streams from the CAFB process (RESOXTM ash and spent stone) contained very little extractable organic material. Organic analyses performed on the spent stone and RESOXTM include HPLC with fluorescence detection to analyze for specific PNA's. The results of this analysis are presented in Table 5-12. The RESOXTM ash will probably be burned to recover its latent heat before it is discarded. During final combustion most of the naphthalene present in the ash is expected to decompose.

TABLE 5-12. RESULTS OF THE HPLC/FLUORESCENCE FOR POLYNUCLEAR AROMATIC HYDROCARBONS IN SPENT STONE AND RESOXTM ASH

PNA	Discharge concentration ($\mu\text{g/g}$)	
	Spent stone	RESOX TM ash
Naphthalene	<120	35,000
Anthracene	110	57
Fluoranthene/pyrene	<600	46,000
Benzo(a)anthracene	<16	10
Benzo(k)fluoranthene	<10	<10
Benzo(a)pyrene/Benzo(e)pyrene	<5.6	<5.6
Indeno(1,2,3-cd)pyrene	<20	<20
Benzo(g,h,i)perylene	<5.6	6,100
Chrysene	<120	<120

^aThese compounds coelute under our experimental conditions.

Bioassay Test Results --

Emission stream samples from the CAFB demonstration plant were tested for biological activity. The bioassay tests were conducted on flue gas samples and solid waste samples. The flue gas samples were collected in the SASS train cyclones, particle filter, and XAD extract. The particulates (particle filter, 1 μm , 3 μm , and 10 μm cyclones catches) were tested in the Ames assay and the rabbit alveolar macrophage (RAM) assay. The Ames assay was also used to test the XAD extract. The spent stone and RESOXTM ash were

tested for health effects using Ames, RAM, and chinese hamster ovary (CHO) assays. The spent stone and RESOXTM ash leachates were tested for ecological effects (freshwater algal assay) and acute static aquatic bioassay (fathead minnow and Daphnia magna).

Results of the biological testing of CAFB samples are summarized in Table 5-13. Ames test results are listed as simply positive/negative (+/-) for mutagenicity in one or more tester strains of Salmonella typhimurium. Data for specific assay parameters are reported for other tests. These results are rated qualitatively according to the response criteria defined in Table 5-Table 5-14.¹⁰

Health effects testing

The Ames test is a Salmonella/microsome mutagenesis assay, which provides a quantitative measure of mutagenic potential of a test substance. The test result is expressed as the number of histidine revertants per plate. A test result is considered positive when a concentration-dependent response is observed in which the level of mutation exceeds two or three times the spontaneous mutation rate (negative control), depending on the specific tester strain, according to the revised protocol discussed in the IERL-Procedures Manual.¹¹

Of the 5 samples tested in the quantitative mutagenesis assay, only the XAD extract induced concentration-dependent mutagenicity. For this sample, positive responses were observed in Salmonella strains TA98 and TA1537, both in the presence and absence of the rat liver metabolic activation system (S-9 mix). In TA98 without metabolic activation, mutagenicity was observed at the lowest concentration tested, 5 microliters/plate. Addition of S-9 activation mix resulted in similar or lower mutagenic specific activity in both TA98 and TA1537. Certain classes of chemicals (e.g., several of the polynuclear aromatic hydrocarbons) require metabolic activation to test positive in the Ames assay. Since addition of S-9 activation mix did not enhance the total mutagenic activity observed, it is reasonable to assume that polynuclear hydrocarbons alone could not be responsible for the mutagenicity. Test results are depicted graphically in Figure 5-4.

TABLE 5-13. SUMMARY OF CAFB BIOASSAY RESULTS

Sample description	Health effects testing			Ecological effects testing		
	Ames response	Rabbit alveolar macrophage assay EC ₅₀	Chinese hamster ovary clonal assay EC ₅₀	Acute static aquatic bioassay		
				Fathead minnow LC ₅₀	Daphnia magna EC ₅₀	Freshwater algae assay EC ₅₀
<u>SASS Train</u>						
Fine particles (filter + 1 µm)	-	900 µg/ml (LOW)				
Coarse particles (3 + 10 µm)	-	>1000 µg/ml (ND)				
XAD extract	+					
<u>Spent Stone</u>						
Stone	-	>1000 µg/ml (ND)	200 µg/ml (LOW)			
Stone leachate				>5.0% (INC) ^a	1.0% (HIGH)	4% (HIGH)
<u>RESOXTM Ash</u>						
Ash	-	>1000 µg/ml (ND)	58 µg/ml (MOD)			
Ash leachate				>10.0% (INC)	>100.0% (ND)	>50% (MOD-LOW)

^aINC - inconclusive because these were the maximum concentrations tested and 50 percent kill was not observed.

TABLE 5-14. RESPONSE RANGES FOR RANKING OF VARIOUS BIOTESTS

Assay	Activity measured	MAD	Response ranges			Not detectable
			High	Moderate	Low	
Health tests						
Ames	Mutagenesis	5 mg/plate or 500 μ l/plate	<0.05 mg or <5 μ l	0.05-0.5 mg or 5-50 μ l	0.05-5 mg or 50-500 μ l	ND at >5 mg or ND at >500 μ l
	Lethality (LC ₅₀)	1,000 μ g/ml or 600 μ l/ml	<10 μ g or <6 μ l	10-100 μ g or 6-60 μ l	100-1,000 μ g or 60-600 μ l	LC ₅₀ >1,000 μ g or LC ₅₀ >600 μ l
Ecological tests						
Algae	Growth inhibition (EC ₅₀)	1,000 mg/l or 100%	<20% or <200 mg	20-75% or 200-750 mg	75-100% or 750-1,000 mg	EC ₅₀ >100% or EC ₅₀ >1,000 mg
Fish	Lethality (LC ₅₀)	1,000 mg/l or 100%	<20% or <200 mg	20-75% or 200-750 mg	75-100% or 750-1,000 mg	LC ₅₀ >100% or LC ₅₀ >1,000 mg
Invertebrate	Lethality (LC ₅₀)	1,000 mg/l or 100%	<20% or <200 mg	20-75% or 200-750 mg	75-100% or 750-1,000 mg	LC ₅₀ >100% or LC ₅₀ >1,000 mg

MAD = Maximum applicable dose (technical limitations).

LC₅₀ = Calculated concentration expected to kill 50% of population.EC₅₀ = Calculated concentration expected to produce effect in 50% of population.

ND = Not detectable.

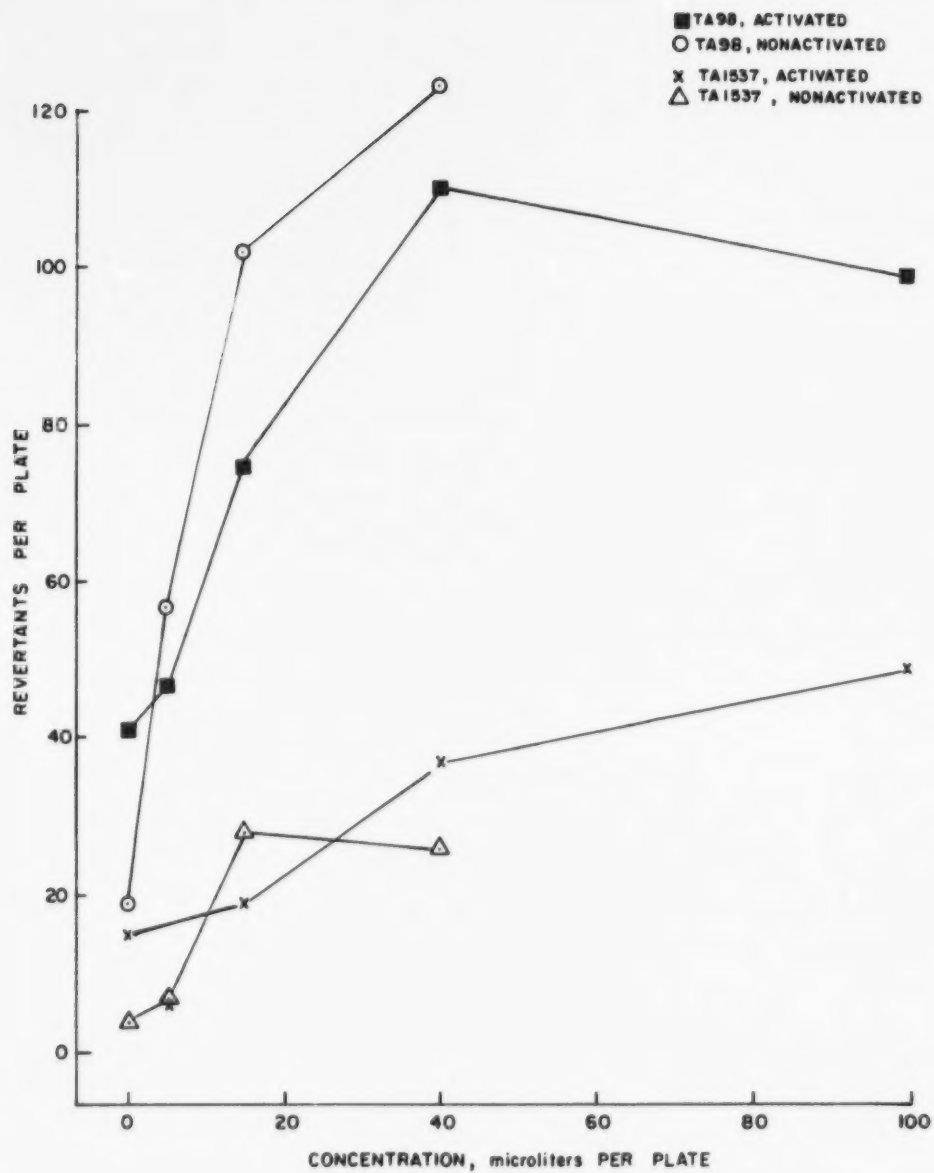


*Figure 5-4. Mutagenic activity of CAFB XAD extract.

In the RAM cytotoxicity assay, three parameters are measured: cell number, cell viability and ATP content. Currently recommended data evaluation criteria require that the toxicity rating of each sample be determined by using the assay parameter with the lowest experimental effective concentration (EC_{50}) value. Based on the condition of this assay and these evaluation criteria, samples of SASS fine particles, SASS coarse particles, and spent stone exhibited a high degree of toxicity to primary cultures of rabbit alveolar macrophage cells in culture. In contrast, the sample of the RESOXTM ash was nontoxic to the RAM cells in culture. These toxicity ratings would apply if the EC_{50} value for ATP content were used (see Table 5-15). However, EC_{50} data for cell number and cell viability reveal slight or no toxicity in all samples. It is unreasonable to assume that intracellular ATP levels could be reduced by 85 percent or more without a corresponding reduction in cell viability. Accordingly, EC_{50} values for cell viability are used to evaluate toxicity of these samples in the RAM test. EC_{50} data for cell viability were the recommended parameter for qualification of cytotoxic response in the RAM test in previous IERL-environmental assessment efforts. Use of cell viability for CAFB samples enables meaningful comparisons to be made between these results and testing of samples from other programs. Laboratory test data can be obtained from the EPA Project Officer.

TABLE 5-15. CAFB RAM TEST RESULTS

Sample	Parameter (EC_{50} μ g/ml)		
	Cell number	Cell viability	ATP
SASS fine particles	1000 (ND) ^a	900 (LOW)	10 (HIGH)
SASS coarse particles	1000 (ND)	1000 (ND)	10 (HIGH)
Spent stone	~1000 (LOW-ND)	1000 (ND)	10 (HIGH)
RESOX TM ash	1000 (ND)	1000 (ND)	1000 (ND)

^aResponse criteria according to Table 5-14.

Results of the CHO clonal cytotoxicity test reveal that spent stone was less toxic to colony survival of CHO cells than was RESOXTM ash. EC₅₀ values were 200 µg/ml and 58 µg/ml, respectively. The EC₅₀ value for the positive control (CdCl₂) was approximately 85 ng/ml.

Based on the conditions of this assay and the evaluation criteria established by EPA, the sample of spent stone exhibited a low degree of toxicity, while the sample of RESOXTM ash exhibited a moderate degree of toxicity to the colony survival of Chinese hamster ovary cells in culture.

Ecological effects testings

The 48-hr median EC₅₀ for the Daphnia exposed to spent stone leachate and RESOXTM ash leachate were 1.01 percent (0.95 confidence interval = 0.85-1.8 percent) and >100.0 percent, respectively. Toxicity of the spent stone leachate is rated as high in this test, while the toxicity of RESOXTM ash leachate is not detectable.

Under the conditions of an Algal assay, Selenastrum capricornutum in the negative control group grew from 1×10^4 cells/ml to 2.33×10^6 cells/ml in 120 hours. Spent stone leachate, when present in the growth medium in concentrations ranging between 0.001 to 0.1 percent did not inhibit the growth of the cells. In fact, these concentrations stimulated the growth of the cells somewhat. However, when the concentration of this test sample reached 1 percent in the growth medium, substantial inhibition (approximately 40 percent), compared with the control, was observed. This growth inhibition remained constant when the concentration of the test sample increased up to 50 percent. The EC₅₀ was calculated to be 4 percent.

RESOXTM ash leachate was not inhibitory to the growth of the algae at the concentrations tested. In fact, a stimulation ranging from 3 percent to 32 percent was observed for these concentrations.

The Freshwater Algae 120-hour screening test provides a quantitative measure of the toxic potential of a test substance to algae growth and reproduction. Spent stone leachate with EC₅₀=40 percent appears to be highly toxic to Selenastrum capricornutum according to the criteria defined in Table 5-14. RESOXTM ash leachate EC₅₀ could only be reported as >50 percent which is moderate-low toxicity.

Conclusions derived from bioassay test results

Based on results of the biological testing of CAFB samples the following conclusions may be presented:

- o SASS particles - No appreciable toxicity in the RAM test; no mutagenicity. No further testing required.
- o XAD extract - Mutagenic in Salmonella tester strains TA98 and TA1537 both with and without metabolic activation. Lack of additional sample precludes any follow-up testing.
- o Spent stone - The solid sample is not mutagenic and has minimal cytotoxicity. Testing of the leachate revealed a highly toxic response to daphnids and algae. Inconclusive results were obtained with fathead minnows. Retesting could be advisable if sufficient sample is available. Presence of the precipitate following leachate neutralization might make the test results suspect.
- o RESOXTM ash - The solid is not mutagenic in Salmonella. This sample demonstrated some toxicity to CHO cells but not toxic response in RAM testing. Testing of leachate showed no appreciable toxicity. Retesting of this sample in the CHO test may be warranted.

Lignite Run 7 - December 1980

The first coal tested was a West Texas lignite, an analysis of which is presented in Table 5-16. Lignite was first introduced to the gasifier at the end of the fifth run. However, an inadequate flow of seal gas to the conveying tables resulted in the termination of the coal feed. The next attempt was during the sixth run. This effort failed at sustained coal gasification because surface moisture on the coal was creating problems with the coal distributing tables, the solids cooler, and other auxiliaries. Successful gasification was realized during the seventh run. For a period of approximately 3.5 hours, lignite was the exclusive fuel for gasification. The seventh run was terminated due to difficulty in transferring damp lignite from the main coal storage tank and due to the inability to dispose of spent fines. The seventh run represents the best lignite run and is discussed in this section.

The sixth run was very short-lived, and as a result, the decision was made to reuse the bed. The seventh run was the first time a sulfated bed was reused. Gasification was initiated on December 9, 1980. Problems with the coal distributing tables necessitated slumping the bed. The bed was refluidized

TABLE 5-16. WEST TEXAS LIGNITE ANALYSIS

Components	As received (%)	Dry (%)
Ash	23.71	30.60
Carbon	40.37	52.10
Nitrogen	0.59	0.76
Hydrogen	1.03	1.34
Sulfur	0.57	0.74
Oxygen	11.21	14.46
Moisture	22.52	---
Higher heating value		
KJ/g	16.01	20.66
Btu/lb	6,885	8,886

on the 10th and on December 11, 1980, oil gasification was started up. Problems with the product-gas cyclone resulted in the bed being slumped again. On December 12, 1980, gasification was again initiated. At 2:48 p.m. the coal feed began. The oil fed to the gasifier was backed out as the coal increased. For 3.5 hours the gasifier operated with lignite as the only fuel.

Averaged values of the emissions data gathered during this period are summarized in Table 5-17. The oxides of nitrogen emissions are lower than the NSPS for lignite-fired power plants of 260 ng/J heat input ($0.60 \text{ lb}/10^6 \text{ Btu}$). Nitrogen oxide emissions obtained during the combustion of natural gas and CAFB product gas are lower than the emissions from the combustion of natural gas alone; emissions during natural gas combustion were on the order of 129 ng/J ($0.31 \text{ lb}/10^6 \text{ Btu}$). Therefore, since natural gas accounted for 35 to 50 percent of the heat input during this test, it is possible that nitrogen oxide emissions would be even lower if lignite derived product gas were the only fuel fed to the boiler.

The sulfur removal efficiency was poor during this run as indicated by the sulfur dioxide emissions presented in Table 5-17. Because of the large quantity of natural gas utilized in this run, the sulfur dioxide emissions are corrected for dilution resulting from natural gas combustion. The corrected values are also presented in Table 5-17. It is clear that the sulfur dioxide emissions are higher than the NSPS for lignite-fired power plants of 520 ng/J ($1.2 \text{ lb}/10^6 \text{ Btu}$).

The cause of low SRE was probably the result of several factors. ERCA has shown that optimal SRE occurred with bed temperatures in the 843°C to 899°C (1550° to 1650°F) range, whereas during this test, bed temperatures were in the 927° to 954°C (1700° to 1750°F) range. In addition, the gasifier bed had a relatively high sulfur content (11.3 percent) prior to initiation of coal feed. It is also noted that the bed sulfur steadily declined throughout the run. This might indicate a release of sulfur from the bed instead of a capture -- perhaps resulting from the higher bed temperature and oxygen stoichiometry stoichiometry (approximately 30 percent of the oxygen required for complete combustion instead of the design rate of about 22 percent). Finally, the regenerator was not in service for the majority of operation, and when it was

TABLE 5-17. AVERAGED EMISSIONS FROM RUN 7 (LIGNITE)

Location	Oxides of Nitrogen		Sulfur Dioxide		Opacity (%)	Oxygen concentration (Vol %)
	Concentration (ppmv)	ng/J	1b/10 ⁶ Btu	Concentration (ppmv)	Corrected (ppmv)	1b/10 ⁶ Btu
Boiler exit	139	119	0.276	257	753	1.31
Induced draft fan exit	107	91.7	0.213	263	770	1.34
Stack						78.0

^aVolume corrected for dilution from natural gas.

^bThis suggests that sulfur was being released.

in operation, it was not operated in an optimal manner. Therefore, the additional fresh material supplied by the regenerator was absent. However, the limestone flow maintained during coal gasification should have had a positive affect on SRE, being almost 2.5 times the stoichiometric amount for 100 percent sulfur removal.

From the outset of coal feed, stack opacity was abnormally high, ranging from 70 to 100 percent. Near the termination of coal feed, opacity showed signs of declining, although not to the point of being satisfactory. This high opacity indicates that particulate clean-up after the boiler will probably be necessary on future units fired by lignite.

Another lignite run, Run 8 was attempted in February 1981. Gasification lasted approximately 4 hours before a high-pressure differential in the fines removal system baghouse necessitated a shut-down.

Bituminous - Run 11 - June 1981

Following the test runs with lignite, a series of tests utilizing a West Virginia bituminous coal were initiated. An analysis of the bituminous coal used in these tests is presented in Table 5-18. The first attempt was in March. This run lasted almost 6 hours on a coal feed. During that time, the boiler fires were clear and strong but the sulfur removal efficiency was only 9-48 percent. The sulfur dioxide emissions ranged from 472 to 859 ng/J (1-2.0 lbs/10⁶ Btu), while the NO_x emissions were 85.9 ng/J (0.2 lbs/10⁶ Btu). Opacity averaged 20 percent. The third attempt in early June 1981 was the longest bituminous run, lasting a total of 12 hours and 33 minutes. Sulfur removal during this run was about 50 percent. The last coal gasification run was on June 22-23, 1981. Bituminous gasification lasted 9 hours and 11 minutes. Initially, the SRE was 71 percent, but by the end of the run, it had deteriorated to 44 percent. This final run (Run 11) is the best representation of bituminous gasification because of its length and its initially high sulfur removal efficiency.

The eleventh run was divided into three sets of data collection periods each representing a stable mode of operation. The first set had all four of the coal feed tables in service and lasted nearly 7 hours. The second set had three tables operational and lasted for 1.5 hours. The third set had only two

TABLE 5-18. WEST VIRGINIA BITUMINOUS ANALYSIS

Components	As received (%)	Dry (%)
Ash	8.32	8.65
Carbon	73.84	76.75
Nitrogen	1.27	1.32
Hydrogen	5.19	5.40
Sulfur	2.11	2.19
Oxygen	5.48	5.69
Moisture	3.79	--
Higher heating value		
KJ/g	31.07	32.29
Btu/lb	13,371	13,897

tables in operation and lasted 45 minutes. Few data were gathered for the third set. The averaged emissions for the first two sets of data are presented in Table 5-19. Values for both the boiler exit and induced draft fan exit are shown.

The bed material from the previous run was reused in Run 11. In addition, the regenerator was not operational. A high fresh stone feed rate was used to offset the loss of stone from the regenerator. These deviations from the designed operation of the CAFB had a negative effect on the emissions during this run.

As seen in Table 5-19, the nitrogen oxide emissions are again lower than during natural gas combustion. Nitrogen oxides emissions were lower than the NSPS for coal fired power plants of 300 ng/J heat input ($0.70 \text{ lb}/10^6 \text{ Btu}$). During this run natural gas provided 55 percent of the heat input. It is likely that nitrogen oxide emissions from bituminous gasification would be lower if the boiler was operating on product gas alone.

Initially, the SRE was about 71 percent. In the fifth hour of gasification, the SRE began to decrease. It then remained steady for about 3 hours at 44 percent before starting a second sharp decrease. The poor SRE experienced as the run progressed could be attributable to a number of factors. The bed material already contained sulfur from the previous run. In addition, the regenerator was not operating. However, the fresh limestone fed to the gasifier should have offset these conditions because the fresh stone feed was sufficient to maintain a Ca:S ratio of 1.8:1. The most probable cause for poor SRE was the high gasifier bed temperature. The bed temperature averaged 1004°C (1840°F). Optimal sulfur removal occurs when the bed temperature is in the $843\text{--}899^\circ\text{C}$ ($1550\text{--}1650^\circ\text{F}$) range. Pilot studies at England indicated that the SRE declines markedly at temperatures higher than 899°C (1650°F).

During the third gasification run, bed temperatures reached 982°C (1800°F) following a gasifier trip. The increased bed temperature corresponded with a sudden drop in the SRE for that run. Bed material samples taken before and after the trip had a somewhat different appearance. The post-trip material appeared to be darker in color and more glazed (very shiny surface) than the pretrip material. Photomicrographs of this material were taken and appear in

TABLE 5-19. AVERAGED EMISSIONS FROM RUN 11 (BITUMINOUS)

Location	Oxides of Nitrogen			Sulfur Dioxide			Carbon Monoxide			Oxygen concentration (Vol %)
	Concentration (ppmv)	ng/J	lb/10 ⁶ Btu	Concentration (ppmv)	ng/J	lb/10 ⁶ Btu	Concentration (ppmv)	ng/J	lb/10 ⁶ Btu	Opacity (%)
With 4 Coal Feed Tables (total operational time: 6 hours, 56 minutes)										
Boiler exit	85.2	82.5	0.192	424	572	1.33	287	169	0.394	4.7
ID fan exit	89.1	86.3	0.201	352	475	1.11	574	339	0.789	6.4
With 3 Coal Feed Tables (total operational time: 1 hour, 34 minutes)										
Boiler exit	83.3	80.8	0.188	827	1,110	2.60	367	216	0.503	4.6
ID fan exit	72.5	70.4	0.164	700	944	2.20	1,130	665	1.55	6.0
									40.5	

Figure 5-5. As can be seen from these photomicrographs, the majority of the surface was smooth, exhibiting very few pores. This phenomenon probably resulted from the increased bed temperature following the trip and contributed to the severe reduction in SRE during the third run and the generally poor SRE experienced during the eleventh run.

Opacity at the induced-draft fan exit was unacceptably high throughout this run. During normal operation when the regenerator is in service, the fresh stone feed rate will be lower, and a subsequent drop in the opacity can be expected. Opacity during lignite gasification was also high, indicating that a post-boiler particulate control device may be required during coal gasification.

SUMMARY AND CONCLUSIONS

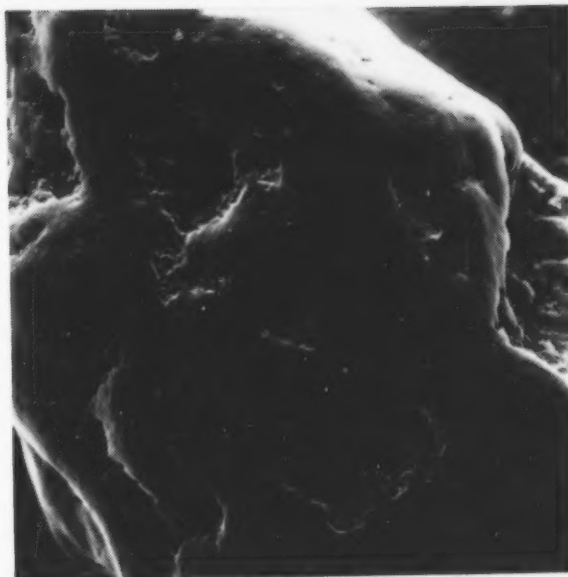
During the 11 test runs, emissions were collected at the boiler exit, the induced draft fan exit, and in some cases, at the stack. Continuous analyzers collected data on sulfur dioxide, oxides of nitrogen, carbon monoxide, carbon dioxide, oxygen, and opacity. In addition, periodic manual emission tests were conducted for particulate, sulfur dioxide, and nitrogen oxides using EPA reference Methods 5, 6, and 7, respectively. A separate program conducted for EPA by GCA/Technology Division characterized the chemical composition of the three major discharge streams from the CAFB: boiler flue gas, spent lime sorbent, and ash from the sulfur recovery system.

In general, the gaseous emissions are below the NSPS for fossil fuel-fired electric generators. The following points summarize the results of environmental interest.

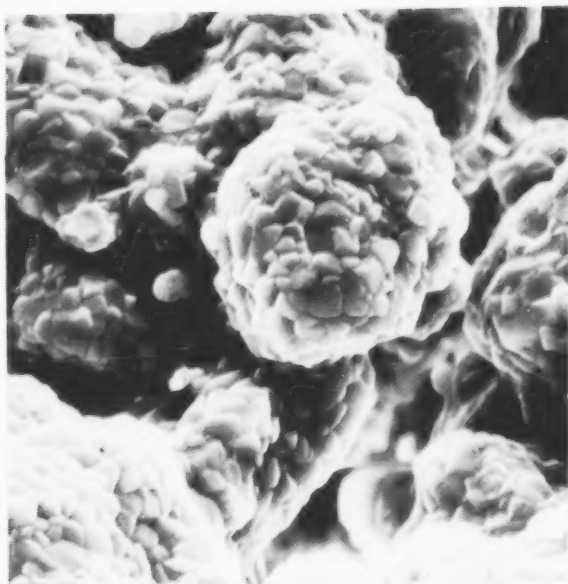
- Due to a higher than anticipated entrainment of bed material, an improved product gas particulate control device is needed. The particulate mass loading exiting the stack was generally within the NSPS for fossil fuel-fired generating units; however, the fines accumulations in the boiler were excessive. These fines were periodically blown out of the boiler causing momentary spikes in the opacity.
- The nitrogen oxide emissions from the combustion of product gas were within the NSPS for electric generating units and were lower than the nitrogen oxide emissions from natural gas combustion.



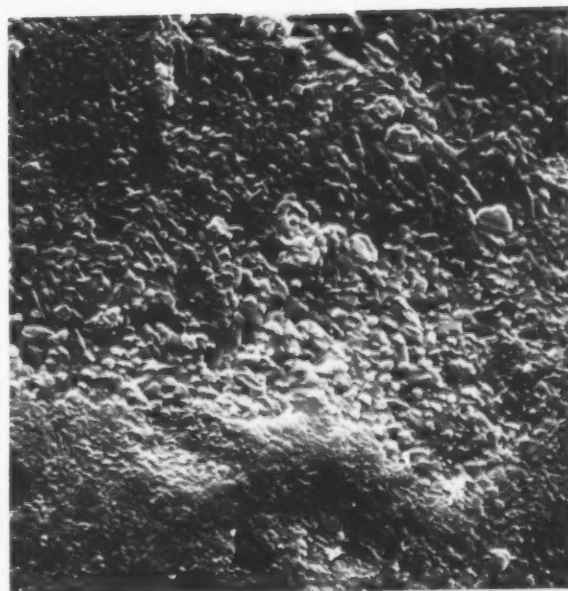
10x



60x



3000x (inside a cavity)



3000x (the smooth surface)

Figure 5-5. Electron Micrographs of bed material after gasifier trip.

- The sulfur removal efficiency (SRE) is greatly dependent on the temperature of the bed. The optimal gasification temperature range is 850°-900°C (1560°F). Regeneration can occur at temperatures above 925°C (1700°C). At high temperatures the bed material becomes glazed (see Figure 5-5) and the SRE decreases markedly.
- Recycling of the ResoxTM tail gas to the gasifier has a detrimental affect on the gasifier SRE, possibly due to the steam and elemental sulfur present in the tail gas.

REFERENCES

1. U.S. Environmental Protection Agency. Code of Federal Regulations. Title 40, Chapter 1, Subchapter C, Part 60, Subpart D - Standards of Performance for Fossil-Fuel Fired Steam Generators. Washington, D.C. Office of the Federal Register. July 1, 1977.
2. Foster Wheeler Energy Corporation. Monthly Technical Progress Narrative No. 56; for EPA Contract No. 68-02-2106. Livingston, New Jersey. March 17, 1980. Run 2.
3. Foster Wheeler Energy Corporation. Monthly Technical Progress Narrative No. 58; for EPA Contract No. 68-02-2106. Livingston, New Jersey. May 19, 1980. Run 3.
4. Foster Wheeler Energy Corporation. Monthly Technical Progress Narrative No. 62; for EPA Contract No. 68-02-2106. Livingston, New Jersey. September 15, 1980. Run 4.
5. Lentzen, D.E., D.E. Wagoner, E.D. Estes, and W.F. Gutknecht. IERL-RTP Procedures Manual: Level 1 Environmental Assessment (Second Edition). Prepared for the U.S. Environmental Protection Agency by Research Triangle Institute, Research Triangle Park, North Carolina. EPA-600/7-78-201, October 1978.
6. Werner, A.S., et al, Preliminary Environmental Assessment of the CAFB. Prepared by GCA/Technology Division, Bedford, Massachusetts, for the U.S. Environmental Protection Agency. EPA-600/7-76-017, October 1976.
7. United States Congress. Resource Conservation and Recovery Act. Public Law 94-580. Washington, D.C. U.S. Government Printing Office. 1978.
8. Environmental Protection Agency. National Interim Primary Drinking Water Regulations. 40 CFR 141 Washington, D.C. U.S. Government Printing Office.
9. Environmental Protection Agency. National Secondary Drinking Water Regulations. 40 CFR 143. Washington, D.C. U.S. Government Printing Office.
10. Brusick, D.J. (Litton Bionetics), C. Hutchinson (Union Carbide), et al, Procedures Manual: Level 1 Environmental Assessment Biological Tests, Draft Report. Prepared by Litton Bionetics, Kensington, Md. and Union Carbide, Terrytown, N.Y. for U.S. Environmental Protection Agency. September 1980.

APPENDIX A

FEEDSTOCKS AND PRELIMINARY TESTS

Feedstocks

The initial shipment of No. 6 fuel oil was received at the job site in November 1978 and the storage tank (390 kl (102,875 gal)) was filled with oil of the analysis as shown in Table A-1.

TABLE A-1. ANALYSIS OF NO. 6 FUEL OIL SUPPLIED TO LA PALMA

API gravity 15.5°C (60°F)	11.90	
SSU viscosity at		
37.8°C (100°F)	4,530	
65.6°C (150°F)	743.5	
98.9°C (210°F)	184.3	
Higher heating value MJ/kg (Btu/lb)	8.80	(18,371)
Ultimate analysis (%)		
Carbon		86.01
Hydrogen		10.56
Oxygen		0.28
Nitrogen		0.32
Sulfur		2.80
Ash		0.02
Moisture		0.01
		100.0
	TOTAL	

The arrangements had been made for the bulk of the fuel oil to be stored by Oil Liquid Terminals Inc., at Port Bownsville. The fuel oil shipment was followed, the same month, by two shipments of limestone from Whites Mines, Texas. These were direct pneumatic discharge truck deliveries with no intermediate bulk storage.

The first was loaded into the limestone storage hopper. Excessive moisture (about 10 percent) was present on the material. The material was cycled around a loop from the bottom of the hopper, through the fill pipe, back to the top of the hopper, in efforts to dry it. This effort was in vain, as the rotary feeder and conveying lines became plugged with wet fines. The limestone was subsequently dumped.

The second shipment of stone was delivered with approximately 2.5 percent moisture, (wet to touch) and was rejected. The supplier was requested to dry the material prior to the next shipment.

The third shipment of limestone was received on December 6, 1978. This load had been dried in a rotary kiln and had an acceptable level of moisture. A sieve analysis of the third shipment was made and is shown in Table A-2.

TABLE A-2. SIEVE ANALYSIS OF THIRD LIMESTONE SHIPMENT

Sieve no.	Sieve size (μm)	Tare weight (g)	Final weight (g)	Through screen (%)	Limestone on screen (g)	On screen (%)
4	4760	535.2	538.9	99.8	3.7	0.2
6	3360	515.9	622.5	94.9	106.6	4.9
16	1190	432.7	1891.3	27.2	1458.6	67.7
30	595	420.0	792.1	9.9	372.1	17.3
50	297	539.5	666.5	4.0	127.0	5.9
100	149	515.1	563.2	1.8	48.1	2.2
140	105	481.0	493.0	1.2	12.0	0.6
PAN	---	464.0	489.3	0.0	25.3	1.2
TOTAL					2153.4	100.0

The size distribution of this stone was similar to that sent to ERCA for testing and also similar to that used by FWEC at Livingston for fluidization and calcination tests. Tests carried out at ERCA indicated that there could be some problems with the initial fluidization of the limestone. Tests at Livingston essentially confirmed this finding, but it was concluded that if the stone was dry, and if there were sufficient fines, then the initial fluidization should be within the capability of the gasifier blower system.

Chemical analysis of a grab sample of limestone is shown in Table A-3.

Several later shipments were delivered and placed on top of the third shipment in the site silo. The size distributions of shipments 4 through 8 are shown in Table A-4. Direct comparison with shipment 3 was not possible because of different cut points.

TABLE A-3. CHEMICAL ANALYSIS OF WHITE MINES LIMESTONE

Component	Concentration
Ca O % wt.	56.0
Mg O % wt.	0.54
Si O ₂ % wt.	2.13
Fe ₂ O ₃ % wt.	0.34
Al ₂ O ₃ % wt.	4.40
CO ₂ % wt.	40.0
Total Sulfur % wt.	0.14
Vanadium ppm	<25
Sodium ppm (as Na ₂ O)	85
Nickel ppm (as Ni ₂ O)	23

TABLE A-4. SIEVE ANALYSIS OF LIMESTONE SHIPMENTS 4 TO 8

Sieve no.	Sieve size (μ m)	% limestone on screen				
		4	5	6	7	8
4	4760	0.4	N.D.	N.D.	N.D.	N.D.
6	3360	4.5	4.4	N.D.	N.D.	N.D.
8	2380	23.2	16.1	21.2	20.8	25.3
10	2000	15.2	13.5	11.3	11.8	10.6
12	1680	14.6	12.7	13.9	13.4	12.3
14	1410	14.6	12.9	14.1	15.0	13.3
16	1190	9.7	9.1	9.9	10.1	9.0
20	841	10.3	13.8	14.0	14.6	13.6
30	595	4.3	N.D.	7.4	7.2	6.3
PAN	---	3.2	17.5	8.2	7.1	9.6

Run-of-mine lignite was arranged to be supplied by the Texas Utilities Generating Corporation (TUGCO) from their Monticello Mine, Texas, to Valley Builders Supply (VBS) who would crush it to <5.0 cm (2 in) store and supply it to the La Palma plant storage as and when required. The lignite deliveries to La Palma were arranged to take place early in the morning so that the morning dew would suppress lignite dust formations. A problem occurred at the Valley Builders Supply when the first load of Texas lignite arrived because VBS's grizzly required that the lignite should be <15.0 cm (6 in) while the fuel from the Monticello mine contained a high percentage of material in excess of that size with consequent plugging of the grizzly. The problem was solved by a switch in the supply to TUGCO's mine in Sulfur Springs, Texas, which produced lignite of acceptable size.

Anthracite for the ResoxTM plant had been procured and was stored by Valley Builders, however, since the ResoxTM process is not described in detail in this report, the analysis and size distribution are not relevant to this report and they are not described.

Cold Fluidization Tests

The mechanical construction was essentially completed by the end of 1978. The refractories were dried and fired in, and most of the electrical construction was completed during January and February 1979. Cold fluidization trials were carried out in the second half of March 1979. These tests confirmed the preliminary findings of ERCA and FWEC in that the available air and flue gas pressure/flow characteristics were sufficient, but only barely, to fluidize the Texas Limestone, and maximum frothing and shroud air to the oil guns was required for the initial lift-off.

A particulate test was performed during the cold fluidization tests and it was determined that while feeding fresh stone at a maximum rate and with the boiler operating at half load, particulate emissions were approximately 25.8 ng/J (0.06 lb/10⁶ Btu). If the boiler was operated at higher load, the emissions would have decreased proportionately.

There were no major problems during the cold fluidization trials, however, flue gas moisture caused some problems, particularly with plugging of the gasifier/regenerator transfer system with damp limestone.

A problem area of future concern was uncovered during this run when investigating a discrepancy in the boiler oxygen reading between two instruments. A six point scan through the flue gave an average value of 4.5 percent with a range of 2.5 to 6.0 percent depending on the sampling location.

During the inspection following the cold fluidization tests, large pieces of limestone were found in several pits in the gasifier distributor. Limestone in the main storage tank was then cycled through a 1.3 cm (0.5 in) screen. Some pieces of limestone captured by the screen as well as some pieces found in the gasifier pits were in excess of 2.5 cm (1 in).

Combustion Trials

Initial Combustion Trial--

The initial combustion trials were carried out mid-May 1979 using a start-up procedure tried out both by ERCA and FWEC. The major steps of start-up were specified as follows:

1. Charge the gasifier with a 46 to 51 cm (18 to 20 in) deep limestone bed.
2. Heat the limestone bed to maximize equilibrium temperature using fluidizing recycled boiler flue gas and twin overhead natural gas burners.
3. During bed heat-up, start other necessary auxiliary systems, e.g., product gas cyclone drains, spent solids transport, line through the lignite feed system, all bleeds and purges.
4. When the bed maximum temperature is reached, feed a sufficient charge of Texas lignite into the gasifier and initiate combustion to enable a temperature of at least 650°C (1200°F) to be reached. If the initial charge is insufficient, supply additional lignite to reach the target temperature.
5. When the bed temperature exceeds the above limit, stop lignite addition and initiate combustion on No. 6 fuel oil with a minimum of six oil injectors.
6. Stabilize the oil combustion and monitor plant performance.
7. Increase the bed depth in the gasifier and initiate stone circulation through the regenerator.
8. Carry out checks of plant performance as necessary throughout the operation and identify/rectify problems.

Bed warm-up procedure (step 2) was initiated on May 12, 1979 and continued through May 13, 1979. The average bed temperature could not be raised above 260°C (500°F) and the pit temperatures could not be raised above 149°C (300°F) despite continuous effort.

In order to raise the bed temperature to the oil ignition point, lignite was introduced into the bed via a manual funnel feed system through two upper coal needle connection points near the ignitor burner on the road side of the unit. With the addition of lignite, bed temperatures in excess of 760°C (1400°F) were achieved and an oil light-off was accomplished on May 22, 1979. During this period, eight oil injectors were in service along the road side of the unit for a period of approximately 20 minutes. During oil combustion operation, the following observations were made:

- The bed pit temperatures on the road side of the unit, were raised to over 760°C (1400°F) and the average bed temperature was raised to over 871°C (1600°F).
- Some above-bed oil flame breakthrough was noted during oil combustion, but this was confined to the top surface of the bed. No freeboard burning or pressure excursions were observed.
- During coal light-off no large clinkers or fusing of bed material was observed.

At this stage sample probes were installed in product gas conduits for sampling gas from the gasifier during start-up. Gas samples analyzed during coal combustion in the gasifier indicated oxygen levels as low as 0.5 percent, carbon monoxide levels as high as 1 percent, and carbon dioxide levels as high as 15 percent (all by volume, dry basis).

Although only the road side oil injectors were in operation, it did appear that bed mixing and temperature propagation were occurring. During the entire operation, the unit appeared to be controllable with no temperature excursions or high-pressure differentials.

The plant was shut down and cooled off.

Problems--

1. During step one of the start-up the limestone feed was limited by the tendency of the fresh limestone to heap at the feed point and neither fluidize nor spread. A temporary solution was to purge the gasifier with air, enter and spread the heap manually. Once there was a layer of fluidized stone in the gasifier the fresh stone spread much more easily provided that it was fed at a fairly low rate, i.e., not more than 360 kg/hr (800 lbs/hr).
2. The Stephens & Adamson reciprocating distributing conveyors caused severe and unacceptable vibrations to the whole CAFB plant and could not be used in their present form. As an interim measure the springs were going to be slackened off and spring mounting adopted as a longer term solution.
3. Lignite spilled from the Barber-Greene weigh belts. Since the skirt boards on the belts were already at their lowest settings, additional idler rollers were needed to prevent further spillage.

During the short combustion run a number of lignite samples were taken at various points in the system to investigate any changes in size distribution. These were grab samples and only approximately representative of the bulk material present at the sampling location. The sieve analyses are given in Table A-5.

TABLE A-5. SIEVE ANALYSES OF LIGNITE GRAB SAMPLES FROM VARIOUS LOCATIONS

Lignite on screen, wt. %					
Sieve size	Sieve (μm)	Storage silo	After delumper	Feed weigh belt	Reciprocating distributor
1/2	12700	8.87	0.0	0.0	0.0
1/4	6350	17.51	4.58	2.13	3.42
4	4760	8.37	4.93	2.93	4.01
6	3360	9.58	8.26	5.01	8.26
8	2380	7.60	8.91	6.95	11.96
12	1680	6.57	8.64	8.32	12.52
16	1190	6.15	8.82	9.96	13.78
20	841	5.56	9.52	11.30	11.67
50	297	14.89	27.03	34.13	22.55
100	149	7.16	12.23	13.07	6.03
140	105	3.78	3.06	5.05	1.54
200	74	1.14	1.43	1.06	1.05
325	44	0.65	0.78	0.36	0.48
PAN	---	2.17	1.81	0.18	2.73
		100.00	100.00	100.00	100.00

Initial Combustion Trial, Second Attempt--

After some necessary repairs and adjustments had been made, early in June another start-up in a combustion mode was attempted using lignite and the ignition burners to bring the bed temperature to the auto ignition point of No. 6 oil, with only about 25 cm (10 in) of limestone bed in the gasifier. The start-up was successful up to the point at which eight of the thirty oil guns were in service in the combustion mode.

Very high overhead and hot gas duct temperatures were experienced and a short time later the metal casing in the annulus region of cyclone CY-101A ruptured. The unit was therefore shut down, cooled and secured for inspection and repairs.

After examining all the available information it was decided that the above bed combustion and the consequent high temperature were caused by low bed height, in the future therefore, start-up bed height should be increased to 61 cm (24 in) if possible or 46 cm (18 in) followed by a rapid build-up if 61 cm (24 in) proves too much for the blowers.

Plant inspection uncovered some damage in the lignite feed system: a sheared drive key in the delumper and torn weigh belts. The coal feed needles were still in very good shape. There was some evidence of erosion where the feed from these needles impacted upon the refractory floor in the

gasifier. This was not entirely unexpected since the velocity through these coal needles during the combustion conditions was deliberately kept very high with maximum flue gas and air to keep the needles cool. Operation in this mode resulted in velocities four to five times the normal velocity. The erosion was repaired with plastic refractory and no further problems were expected in this area. Some slag formation was found in the ducts leading to the cyclones. There was evidence that this slag had been molten because drops and globules were formed from the ceiling. The deposition was not excessive and appeared to have been self-limiting.

First Successful Combustion Tests--

Preparations--Cyclone-101A was repaired and additional thermocouples were added in the annulus area to indicate the skin temperature. The skin temperature would be limited in future, if possible, to 399°C (750°F).

The oil system was changed from a return flow to a dead ended header and the thermal flow meter was replaced by a positive displacement meter. The main oil pump was consequently reduced in output from 416 to 151 l/min (110 to 40 gal/min) and one of the control valves was moved to the pump by-pass line. This arrangement meant that there would be no need to change the process control logic. A sampling point on the gasifier plenum gave an oxygen reading in the control room so that the stoichiometry of the operation could be calculated easily and accurately.

Anti-vibration pads were fitted to the coal feed system to cut down the vibration in the remaining plant. Spring tension on the distributor tables was increased to give 2720 kg/hr (6000 lb/hr) per table and the actual rate was checked by discharging lignite into drums and weighing after a fixed interval of time. The damaged weigh belts were replaced and the read-out heads adjusted to read correctly.

During this stage of operation several lignite samples were taken from various locations in the lignite feed system to check for any product degradation and also to identify any points where fines segregation may be occurring. These data are similar to that in Table A-5 but there is much more of it. Since this size distribution data makes no contribution to this report it is not summarized here but is available in the FWEC Monthly Technical Process Narrative No. 52, September 24 to October 28, 1979, which is available from the EPA Project Officer.

In October 1979 there were three attempts to initiate combustion in the CAFB gasifier: the first two were aborted part way through while the third was successful. The lignite feed system was set up to deliver approximately 2722 kg/table (6000 lb/table) (design rate) by retensioning the springs. The calibration was done by timing a known weight of lignite through the system. The weirs were set so that hold ups on the tables would not occur. Thus it was decided to accept the excessive vibration for the brief period (15 to 20 minutes) that the system would be operated during start-up. The coal downcomers and needles were re-fitted.

The usual system checks were carried out--no problems encountered. The boiler was in service and stabilized at 50 percent load (10 MW).

Start-up No. 1 (October 5th)--The limestone bed was fluidized with maximum boiler flue gas recycle plus 5 percent air and with maximum overbed heating using the natural gas burners. Stone feed was started to build up the bed to 46 to 51 cm (18 to 20 in) after the product gas cyclone drains were warmed up by venting via the rodding points only, and with flue gas circulation through the solids transport system to the solids cooler.

The limestone feed rate was limited to approximately 364 kg/hr (800 lb/hr) due to poor fluidization at low temperature, and there were frequent interruptions in the feed to enable accumulated mounds of stone below the inlet to disperse.

By October 5th, the bed depth was approximately 45.7 cm (18 in), and the temperature had stabilized at approximately 220°C (430°F). The cyclone drains were clear and all systems were functional.

Lignite was added at maximum rate via all feeders for 15 minutes. It was estimated that 2700 kg (6000 lbs) had been added representing approximately 25 percent of the bed weight. (Similar to the proportion used during ERCA start-ups.)

The bed was slumped for 20 minutes and on refluidization the temperature increased smoothly but reached only 325°C (620°F). At this point, the start-up could have proceeded on lignite combustion, but it was found that all the needles were plugged with lignite. This probably occurred when the lignite feed system was shut down and the bed was slumped. These steps were taken simultaneously but some lignite could have been carried off the tables while the gasifier bed was slumped. It was concluded subsequently from the small temperature rise that there were hold-ups in the lignite feed system at this time, other than the needle plugging.

Start-up No. 2 (October 5th)--This was a repeat of the above having cleared the lignite needles except that the lignite feed was run for 40 minutes, and the bed was not slumped since there were signs that combustion was already in progress during the feeding period. Fluidizing air was increased at the expense of flue gas and the bed temperature increased smoothly over a period of about 45 minutes to 730°C (1350°F). Fuel oil injection was started on six guns when the lignite was burned out.

Several problems occurred simultaneously. There was considerable overbed burning, the cyclone temperatures increased rapidly and within 30 minutes or so had reached over 540°C (1000°F) and the boiler flame (natural gas) became unstable. In view of the previous cyclone burn-out, the start-up was aborted at this stage and the plant was shut down.

On subsequent investigation it was found that the plant had been operating in a "lean gasification" mode. Gasifier stoichiometry was approximately 80 percent with the oil injection pit stoichiometry at

27 percent. The overhead burning was aggravated by the further combustion of the product gas by excess air from the natural gas warm-up burners. In turn this escalated the cyclone temperatures and the heat input to the boiler caused the temporary flame instabilities.

After shut down the gasifier bed was cooled, initially with air but later with flue gas. Unfortunately full flue gas flow was not maintained through the coal needles. This had been "stopped" by closing the flow control damper but it was subsequently found that due to the flue gas recycle or pressure, the minimum flow through this damper was still 900 kg/hr (2000 lb/hr) or 40 percent of maximum. This was not sufficient to keep the coal needles clear as the temperature of the bed dropped, and also to prevent condensation in the distributing tables and downcomers. As a result, all the coal needles were found to be plugged with limestone plus hydrated, partially calcined lime. In most instances, this extended up into the downcomers. All 24 feed lines had to be dismantled and drilled out.

The cyclone internals were inspected and no damage was found.

Additional serious problems were found in the lignite feed system at this time. On both A and D tables, plugging at the table outlets had caused backing up of the lignite, flooding the tables themselves and backing up through the rotary feeders, and the metering weigh feeders so that both weigh feeders were unusable. The belts were torn and the weigh feeder housings were packed with lignite in both cases.

Also it was found in C table that the lignite was smoldering. As it could not be removed the table was resealed so that the risk of fire would be minimized by the self-inerting effect of the smoldering. (A close watch on the temperature in this table showed that this did indeed happen.)

While replacement parts were being obtained for tables A and D and repairs were carried out, an attempt was made to start the plant using feed tables B and C only.

Start-up No. 3 (October 12th)--The gasifier bed depth was increased to 51 cm (20 in) while the vessel was open during the cyclone inspection.

The procedure adopted was as previously described except that the addition of lignite was made using only B and C tables running at 50 percent of capacity for 1.25 hours. This limited the chance of needle blockage and allowed for the lignite consumed during the feeding period. During this operation, three needle blockages were detected and cleared.

Without slumping, the bed temperature rose uniformly to 815°C (1500°F) showing that excellent bed mixing was achieved during the lignite addition. The switch to oil injection was easily achieved, with six oil guns in service and the bed temperature rapidly stabilized at approximately 790°C (1450°F).

It was, however, almost immediately apparent that the cyclone temperatures were again increasing rapidly and actions were taken to limit this. First, the overhead natural gas was stopped but the air supply continued at maximum rate as a coolant. Second, the oil distribution to the bed was improved by putting more oil guns in service but at the same total oil flowrate. This required that the pressure drop across the metering orifices be reduced below the recommended minimum 34.5 kPa (5 psi) and this in fact was reduced to 13.8 to 20.1 kPa (2 to 3 psi) with eleven guns in service without encountering any oil distribution problems. Other actions taken but probably of lesser influence were to reduce the amount of gas flow through the coal needles and to reduce the fluidizing gas flow while maintaining good bed mixing.

The cyclone skin maximum temperature in the critical annulus regions peaked at 604°C (1120°F) and was eventually stabilized at 582°C (1080°F). However, with the gasifier temperature at only 790°C (1450°F) it was obvious that the cyclone cooling fins were ineffective and the skin temperatures were far in excess of design (cyclone design was for a skin temperature of 316°C (600°F) with a gas temperature of 1370°C (2500°F)). This effectively prevented any increase in the gasifier bed temperature beyond about 790°C (1450°F).

The plant was therefore shut down after 66 hours of steady combustion.

Valuable experience had been gained in operating the plant and equipment and some salient points are listed here since they were generally applicable to subsequent tests. (Note: wherever possible a comparison is made with similar experience on the ERCA pilot plant in England.)

1. The initial combustion was at one end of the gasifier only but it spread rapidly throughout the gasifier bed. Thereafter the gasifier bed temperatures including the pit temperatures, were very uniform showing that bed mixing was excellent. It was usually not possible to detect from temperature alone which pits were receiving oil.
2. It was possible to switch oil injection pits at will provided a reasonable distribution of oil was maintained across the gasifier.
3. At no time, including start up on lignite or stone addition, did any difficulties occur with the cyclone drain system.
4. All skin temperatures except for the cyclones were satisfactorily low.
5. Stone feed at a rate up to approximately 360 kg/hr (800 lb/hr) was possible without difficulty. This would, under normal circumstances be more than adequate. However, with the limitations on temperature and velocity, higher feed rates were not always possible because they resulted in occasional loss of pit temperatures local to the stone entry. These excursions were easily handled--by reducing stone feed rate and switching oil nozzles. They did not prevent operation of the stone feed system at maximum rate (1360 kg/hr or 3000 lb/hr) for short periods (hours).

6. At the typical bed velocity of approximately 0.58 m/sec (1.9 ft/sec) there was no evidence of defluidization.
7. During the combustion, typically the overall gasifier stoichiometry was 180 percent with the oil pit stoichiometry at about 55 percent.
8. There was no apparent difficulty in boiler operations and fines accumulation in the boiler did not appear to cause any serious impairment of efficiency.
9. Viewed from above, the gasifier bed appeared very similar to the ERCA unit (uniform orange/red, occasional licks of flame, under combusting conditions). Thus, the somewhat alarming appearance of the gasifier bed overburning when viewed from the side may be typical and need not be of major concern.
10. All the rodding operations carried out, on the lignite needles, the oils guns, the solids transfer system, and the gasifier pressure tappings presented no problem whatsoever. There was no excessive ejection of hot solids, and if anything rodding out was a lot easier and less hazardous than at ERCA because of the ease of access to the rodding locations.
11. Data collection was excellent via the data logger. This system also had facility to print the visual screen display and to display curves of individual plant parameters over a period of up to 4 hours.

However, there was no display of major plant variables, updated at regular intervals (as there was at ERCA), which could have made easier monitoring of plant conditions.

There were a number of problems with equipment which required attention or modification:

1. Oil nozzles were rigidly connected and could not be removed (two nozzles were plugged while another six were not available because of an obstruction in the line, valve or orifice) while the plant was in operation and therefore could not be cleared or changed until shut down.
2. There was dust leakage through the glands on the Ducon rotary valves at the bottom of the cyclone drain legs.
3. Pressure tappings required frequent rodding.
4. The gasifier bed depth was not high enough to allow rapid transfer of bed between the gasifier and the regenerator, other than by splash. However, the transfer system became plugged with damp partly calcined limestone quite early in the start-up and so would have been inoperable. The pulsing system (working on flue gas) filled with condensate and, for some of the time, the solenoid valves were not operational.

5. The lignite distributing tables were capable of causing the whole CAFB plant to vibrate, and this was severe when the tables were operating at their maximum throughput (gasification rate). This was particularly true of C and D tables which did not have the benefit of adjacent damping masses as did A and B tables which are located near the massive product gas cyclones.

Damping devices had been fitted but a pronounced vibration problem still existed at high feed rates.

The maximum feed rate of the vibrating tables is governed by adjustment of the return springs tension. Thus the vibration problem would be reduced but only at the expense of feed rate which meant essentially that gasification could not be achieved.

Changes Indicated--

Procedural--

1. A minimum gas flow velocity of 9 m/sec (30 ft/sec) needs to be maintained through the coal needles at all times when the bed is fluidized. When the bed is not fluidized then the gas flow to the coal needles needs to be positively shut off--not just by closing the control damper.
2. Lignite feed should be stopped if possible by shutting off the weigh feeders and the rotary valves and running the tables empty (approximately 10 minutes) into the fluidized bed, before shutting down.

Mechanical--A positive cooling system, either by water spray or blown air, needs to be installed on the cyclone outlets to prevent excessive skin temperature.

Results

Since this was essentially a plant test-out, little formal data collection was made, however, the opportunity was taken to try out the sampling and measuring systems. In particular, gas samples were extracted from the centroid of the conduit ahead of cyclone CY 101B and O₂, CO₂, CO, NO_x and opacity data taken at least once an hour. A summary of the product gas analysis data is shown in Table A-6: the first data line shows the conditions during early start-up when No. 6 fuel oil was first injected while there was still an inventory of devolatilized lignite in the gasifier bed. Oil and lignite partly combusted partly gasified in the bed; combustion was completed above the bed and all, or nearly all oxygen was used up. The bed temperature at this point was well in excess of 788°C (1450°F) and rapid decomposition of limestone set in so that the CO₂ increased well above the stoichiometric level.

The second line of data in Table A-6 refers to No. 6 fuel oil combustion with air and no flue gas to the plenum, towards the end of the run when the bed temperature was being reduced gradually by lean mixture combustion.

The third line of data in Table A-6 is an average of the bulk of the collected data but excluding the extremes represented by the two lines above.

TABLE A-6. PRODUCT GAS ANALYSES: SAMPLE POINT AHEAD OF CY-101B

Condition	% by volume		Opacity 1 pass, %	ppm by volume		
	O ₂	CO ₂		CO	SO ₂	NO _x
Start	0.45	>20	100	>20000*	70	20
End, lean combustion	15.2	3.9	0	0	70	87
Run average	12.4	6.9	10	420	70	72

*Note: instrument off scale

Gasifier and regenerator bed samples were taken: gasifier samples from the bottom of the front and rear beds, and also from the top of the bed at the end opposite the regenerator. The regenerator was emptied during the combustion and some material was passed across from the gasifier to the regenerator during the early attempts to achieve transfer. Samples of the regenerator bed surprisingly exhibited similar analyses to those from the gasifier bed. The results of the analyses are shown in Table A-7.

TABLE A-7. COMPOSITION OF BED MATERIALS (wt %)

Sample from	Total S	Total C	Carbonate C
Gasifier, front bottom	7.5	4.6	0.24
Gasifier, rear bottom	7.5	2.0	0.24
Gasifier, middle top	7.6	N.D.	N.D.
Regenerator top	5.4	N.D.	N.D.
Regenerator middle	6.1	4.8	0

APPENDIX B

PART 1: FUEL OIL GASIFICATION (RUN 1)

Preparation

A cyclone cooling system was installed. This system consisted of an air blower and a manifold which directed cooling air to the base of the cooling fins on the Emtrol product gas and regenerator cyclones. Flexible hoses were added to both air and oil connections on the 30 gasifier oil guns and also on the the two stand-by guns in the regenerator, to permit on-line removal for cleaning plugged injectors.

Startup on Combustion

The gasifier containing 51 cm (20 in) of stone was preheated with the overhead burners and fluidized with hot flue gas on December 16, 1979. Lignite was injected at about 19:30 and some air admitted to the plenum. By 19:55 the gasifier temperature started rising and reached 399°C (750°F) but did not go much higher. At 21:00 lignite feed was discontinued (about 3,630 kg or 8,000 lb lignite had been fed in) and at 21:15 the beds were slumped. The gasifier was fluidized at 22:20 with immediate ignition: bed temperatures started rising very rapidly, duct O₂ dropped to 0.2 percent, and CO went off scale on the meter (max reading 2 percent). Pit temperatures reached 704°C (1300°F) by 22:40 and at 22:46 eight oil guns were put into service. Combustion of No. 6 fuel oil was maintained until noon of December 19 when gasification was initiated.

During the 3 days on combustion the operating crews experienced the following problems and new operations:

- (1) Low fluidizing velocity trip.
- (2) Blockages in cyclone drains.
- (3) Trips on cyclone drain rotary valves.
- (4) Blockage of fines lines to spent solids cooler.
- (5) Tried out stone feed at 100 percent: indicated 1,420 kg/hr (3,132 lb/hr).

- (6) Removed and replaced oil guns.
- (7) Shut down for 6 hours to reset cyclone drain rotary valve rotor clearances, also to clear out cyclone drains and clear solids line to the spent solids cooler. Restarted with no problems.
- (8) Found flue gas baghouse discharge conveyor plugged and bridged over with hot fines which had an angle of response of over 90°.
- (9) Varied the number of oil guns in the range 8 to 14: it was more convenient to control combustion temperature by switching oil guns in and out of service at a constant oil manifold pressure than to control the total flow of oil to the unit with a fixed number of guns in service.
- (10) An attempt to start the regenerator was made by placing coal on the bed surface, heating with the startup burner and combusting. This effort was unsuccessful as the coal burned on the surface causing considerable agglomeration of ash and bed material, forming a crust on the surface. This made it impossible to fluidize the regenerator bed.

In preparation for gasification:

- (1) All the oil guns were test fired and cleared if necessary (28 O.K., 2 cleared).
- (2) Igniter burners were lit in burners 3 and 4.
- (3) Temporary damper was rigged to control the air flow through the gasifier air blower B-101. It was found that clamping a steel plate over most of the inlet silencer top flange allowed a much finer control over the gasifier air flow at very low air rates. This was necessary because the main air damper leaked a considerable amount of air at shut off.

By about 10:10 in the morning of December 19, 1979 the product gas oxygen had been reduced to 0 to 2 percent, at a gasifier temperature of about 840°C (1550°F), as a result of backing off gasifier air to a trickle, fluidizing with flue gas and combusting the oil (about 450 kg/hr (1,000 lb/hr)) nearly stoichiometrically: mainly with frothing air and air contaminating the flue gas. Boiler flue gas oxygen was steady at 6.5 percent.

The Changeover (Noon December 19, 1979)

The conditions set in the morning were maintained with only small changes until noon at which point the conditions were:

- Air and flue gas to plenum 9,100 kg/hr (20,000 lb/hr)
- Oil rate 523 kg/hr (1,150 lb/hr)

- Temperature 877°C (1610°F)
- Product gas O₂ 0.7 percent by vol.
- Overhead burners out of service, no air flow
- Eight oil guns in service.

The actual changeover followed in quick succession as follows in Table B-1.

At this point the gasifier temperature was 871°C (1600°F), after passing through a maximum of 927°C (1700°F) with product gas burning in both burners (No. 3 still without a pilot) and generating about 4 MW on product gas, both furnace and stack were looking clear. No. 6 rear oil gun was plugged up but was soon cleared and put into service, thus by 12:53 there was 28 guns in service and a consolidation period began. The total boiler load at this point was 10 MW, 6 MW being contributed by natural gas.

12:43 - 16:00 The oil rate was maintained constant at about 1,905 kg/hr (4,200 lb/hr), total air and flue gas flow to the plenum was increased to 9,980 kg/hr (22,000 lb/hr) by increasing the air rate, and the limestone rate which was at 408 kg/hr (900 lb/hr) was increased to 726 kg/hr (1,600 lb/hr). The changes were made at the beginning of this period and the conditions were maintained constant. The gasifier temperature dropped to a little above 820°C (1500°F) and remained approximately constant, rising very slowly.

16:00 - 19:00 Over the next 3 hours the oil rate was increased and natural gas backed off in several increments to 3,310 kg/hr (7,300 lb/hr) starting with a slow rise up to 2,114 kg/hr (4,660 lb/hr) by 17:00 and a rapid increase to 2,345 kg/hr (5,170 lb/hr) at about 17:00. At the same time the temporary throttle on the inlet to B-101 air blower was removed. The next increase was to 2,850 kg/hr (6,280 lb/hr) at about 17:40, followed by a final increase to 3,311 kg/hr (7,300 lb/hr) at about 18:30. At this point the unit was running at 10 MW of which 9.7 was supplied by the CAFB and about 0.3 by the pilot on No. 4 burner. The total air and flue gas flow to the plenum was increased to 12,200 kg/hr (27,000 lb/hr) during that time. The gasifier temperature, which was 838°C (1540°F) at 16:00 gradually increased to 882°C (1620°F) at 17:30 and thereafter remained approximately constant at 882°C (1620°F).

Steady Run

From 19:00 on December 19, until about noon on December 20, the gasification conditions were maintained as near as possible constant, at:

- 28 guns in service
- 3,310 kg/hr (7,300 lb/hr) No. 6 oil rate
- 12,250 kg/hr (27,000 lb/hr) air and flue gas to plenum

TABLE B-1. COMBUSTION TO GASIFICATION CHANGEOVER SEQUENCE

Time	No. of oil guns in service	Oil rate kg/hr (lb/hr)	Percent stoichiometry	Notes
12:00	8	522 (1150)	103	Combustion
12:03	10	592 (1306)	91	Gasification
12:04	12	670 (1476)	80	Gasification
12:05	12	719 (1586)	75	Gasification
12:08	12	719 (1586)	75	Trace flame on burner No. 4
12:09	14	773 (1705)	69	Trouble with No. 3 igniter
12:10	18	1149 (2534)	47	
12:15	18	Increasing to 1419 (3128)	38	No. 4 burning well No. 3 unsteady
12:20	18	1520 (3352)	35	
12:23	22	1548 (3412)	35	
12:35	27	1978 (4360)	27	

- 580 kg/hr (1,280 lb/hr) limestone feed rate
- 900°C (1650°F) gasifier temperature

There were some changes caused by a trip out of the limestone feed system (out 19:00 to 20:10 on December 19 and 8:50 to 10:00 on December 20) and plugging up of 6 rear and 1 front oil guns: both of the oil guns were taken out of service, cleared and put back into service. Before noon the temporary damper was replaced in nearly open position on the 8-101 gasifier air blower inlet in preparation for sulfation and shut-down, and the air and flue gas to plenum was reduced to 11,340 kg/hr (25,000 lb/hr).

Burner and Oil Gun Tests

At 14:00 an attempt was made to improve the product gas flame pattern by restricting the air register of the No. 3 product gas burner and by removing the pilot gas from the No. 4 burner. The air register was reopened since it seemed to have no effect, but removal of the pilot improved the flame. Still there was some flame impingement on the side wall and on the superheater tubes. At 17:30 an experiment was started on reduction of the oil guns in service starting with 28 guns, i.e., all with the exception of 15 rear and 13 front (see Figure B-1 for oil gun nomenclature). The sequence followed was as shown in Table B-2.

At this point there were eleven guns in service, 2, 3, 6, 9 and 12 front and 2, 3, 6, 9, 12, and 14 rear. The oil flow was adjusted back to 3,310 kg/hr (7,300 lb/hr) after each change, however after the penultimate change (to 10 guns) the oil control valve was found to be wide open and there was a reduction in flow which necessitated one gun (3 rear) being brought back into service. During these tests the only noticeable change was that in boiler SO₂ which remained constant at about 100 ppm until 18:00 then rose to 150 ppm at 18:16 then rose sharply to 190 ppm and dropped back. At 19:00 guns 7 front, 7 rear and 14 front were brought back into service (i.e., 14 guns in all) and the boiler SO₂ dropped back to 130 ppm and leveled out. Throughout these tests there was no noticeable difference in the gasifier temperature, pit temperatures or boiler flames.

Conclusions from this test are summarized below:

m ² Bed area/ oil gun (ft ²)	Effect on	
	Gasification	Desulfurization
0.74 or less (8)	None	None
1.0 (11)	None	Reduced by 3%
1.3 (14)	None	Reduced by 5%
1.4 (15)	None	Reduced by 8%

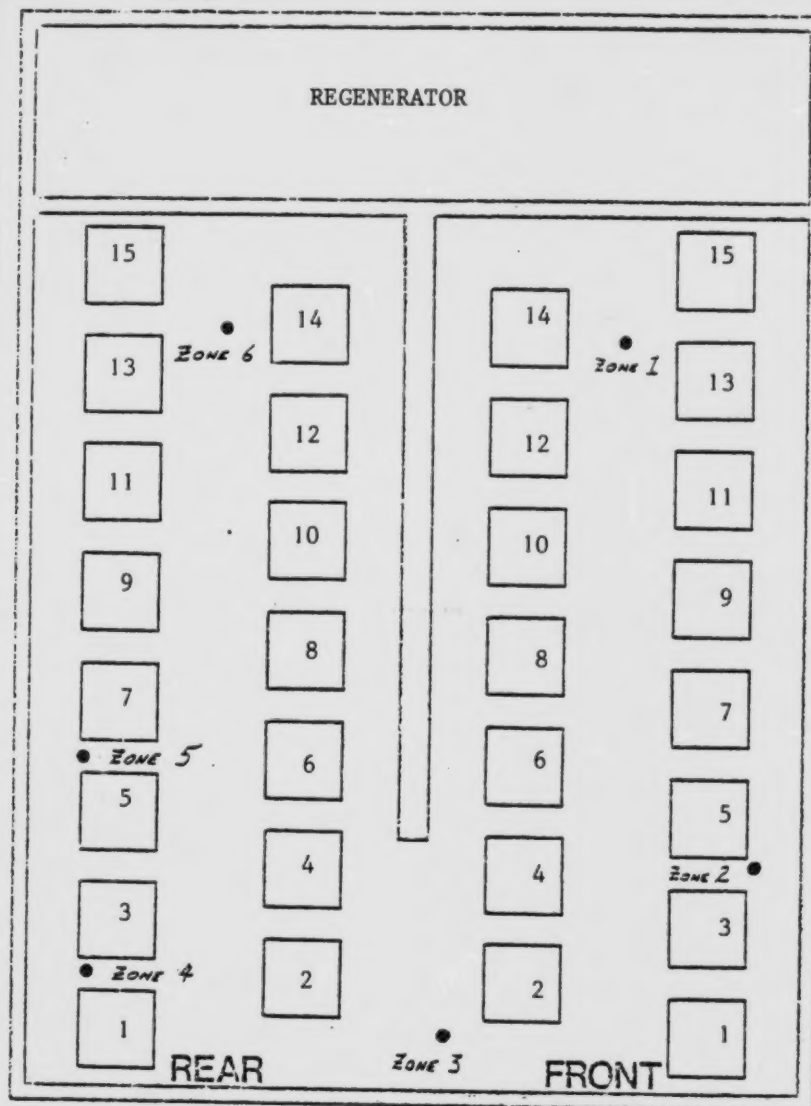


Figure B-1. Pit lay-out and gasifier thermocouple locations.

TABLE B-2. SEQUENCE OF OIL GUN CHANGES

Time	Guns removed	No. of guns left in service	Boiler SO ₂ ppm (after change)
17:00	--	28 (start)	100
17:31	1F, 1R	26	100
17:32	15F, 13R	24	100
17:33	4F, 4R	22	100
17:47	7F, 7R	20	100
18:00	10F, 10R	18	100
18:10	8F, 8R	16	120
18:16	11F, 11R, 5F, 5R	12	150
18:38	14F, 3R	10	190
18:39	3R back in service	11	130
19:00	7F, 7R, and 14F back in service	14	130

Changeover From CAFB to Natural Gas and Sulfation

At 19:14 on December 20, igniters in burners 3 and 4 were put in service. At 19:16 oil flow was reduced to 2,940 kg/hr (6,490 lb/hr). No. 1 natural gas burner was put into service and the air and flue gas flow to the plenum was reduced. Over the next hour the oil flow to the gasifier was progressively reduced and natural gas flow increased maintaining a 9MW load on the boiler. Gasifier air was also backed out to keep the gasifier temperature constant at about 900°C (1650°F) or below, by using the air control valve and the temporary blank.

By 20:07 the oil flow was down to 1,070 kg/hr (2,360 lb/hr) through eight oil guns (2, 6, 9 and 14 front and rear) with flame in both No. 3 and No. 4 burners and the plant was ready for transition from gasification to stoichiometric combustion.

At 20:09 the oil flow was reduced to 394 kg/hr (869 lb/hr) and all air was isolated from the main air blower B-101. Gasifier temperature at this stage was 871°C (1600°F) and rising fairly rapidly. The boiler was clear with pilots only in burners No. 3 and 4 and 8.5 MW of natural gas on burner No. 1.

Sulfation

At 20:14 all the oil guns were switched off but frothing air was left on to clear the guns.

At 20:33 all the oil gun rod outs were pushed home and fronting air was valved off. At this point the gasifier temperature had peaked at 932°C (1710°F) as sulfation commenced.

At 20:52 the gasifier air was increased a little giving a gas duct CO₂ reading of 7.5 percent and a temperature which dropped to 887°C (1630°F) rose to 932°C (1710°F) and remained at that level until 23:00 when it started dropping. At 23:28 the duct CO₂ reading had dropped to near zero, oxygen started rising and the bed temperature dropped to 843°C (1550°F), i.e., the sulfation and burn-out were complete. Air and flue gas flow was maintained until 1:30 on December 21, to cool down the bed, and the plant was then shut down.

Results

Throughout the gasification run operating data were collected by the microcomputer data logger, on control room charts, and from samples collected to monitor emissions. In addition, an EPA Method 5 particulate emission test was performed, towards the end of the gasification run, to determine the particulate mass loading.

Contained in Table B-3 is a summary of emission data recorded during gasification. The highest, lowest, and average values are shown for two periods of time:

TABLE B-3. EMISSION RESULTS DURING GASIFICATION

Hours into gasification	Gas or pollutant	Value observed			Mass rate of emission ng/J (lb/10 ⁶ Btu)		
		Highest	Lowest	Average	Highest	Lowest	Average
0 to 6	NO _x , ppm	160	64.0	136.2	177 (0.411)	57.5 (0.134)	109 (0.253)
6 to 30	NO _x , ppm	100	84.0	89.3	66.7 (0.155)	49.1 (0.114)	56.8 (0.132)
0 to 6	SO ₂ , ppm	20	0	6.25	18.5 (0.043)	0 (0)	6.97 (0.0162)
6 to 30	SO ₂ , ppm	135	0	57.1	120 (0.279)	0 (0)	50.3 (0.117)
0 to 6	O ₂ , %	13.5	4.5	8.41	---	---	---
6 to 30	O ₂ , %	6.5	4.2	5.2	---	---	---
0 to 6	CO, ppm	70	0	---	---	---	---
6 to 30	CO, ppm	100	<10	---	---	---	---
0 to 6	CO ₂ , %	11.4	5.4	9.46	---	---	---
6 to 30	CO ₂ , %	13.8	10.1	11.43	---	---	---
0 to 6	Opacity, %	22	15	18.25	---	---	---
6 to 30	Single Pass, %	29	8	15.07	---	---	---

- The initial 6 hours of operation are from 12:00 to 18:00 on December 19. During most of that time 4 to 5 MW was supplied by product gas, the remainder was supplied by natural gas.
- The next 24 hours of operation from 18:00 on the 19th to 18:00 on December 20, 1979 was when the boiler was at 9 to 10 MW load of which all but a natural gas pilot light, was supplied by the CAFB product gas; for a few hours even the pilot was extinguished.

Data for the final 2 hours of gasification are not shown as this was a period of transition between gasification, natural gas combustion and sulfation. Throughout the operation on product gas only (hours 6 through 30) NO_x and SO_2 emissions never exceeded the NSPS limits for gaseous or liquid fossil fuel utility boilers even though this period included most of the tests on the number of oil guns in service.

During the same 24 hour period, the opacity standard (20 percent) was exceeded for 4.25 hours at the beginning of this 25 hour period but never exceeded this limit after that.

The EPA Method 5 particulate test indicated a mass loading of 697 ng/J (1.62 lb/10⁶ Btu). This test was performed from 16:00 to 18:00 on December 20, 1979. During this time, opacity varied from 10 to 22 percent. Throughout this test, fresh limestone was fed to the gasifier at a rate of approximately 545 kg/hr (1,200 lb/hr).

Several steady periods during which there were no major changes in input (except for the trip out of the limestone feed system, which did not have a great effect other than a small increase in the gasifier temperature) were chosen, and a selection of the average data for these periods is reported in Table B-2. These data were not corrected for any departures in measured flow (determined later--during Run 4) or any other noted or known deviations. Neither was there any attempt to reconcile the data via mass balances. The only readily discernable trends in Table B-4 are a steady increase in the stack SO_2 level and a decrease in the opacity. Indeed, the trend continued and by the time oil gun trials had started the stack SO_2 had risen to 100 ppm and the opacity had fallen to 15 percent. Samples of gasifier bed material were taken at various points in the test run and their analyses are reported in Table B-5.

It is apparent that the sulfur content of the gasifier bed increased steadily throughout gasification, peaked after approximately 24 hours, and then declined slightly near the end of gasification and during sulfation. When transition from gasification to sulfation was passed through (0 percent stoichiometry), the gasifier bed regenerated briefly (evidenced by SO_2 concentrations in excess of 500 ppm).

During the 26th hour of gasification (14:00 on December 20) samples of fines were extracted from the cyclone dip legs and analyzed with the results as shown in Table B-6.

TABLE B-4. EXTRACTS FROM RUN 1 DATA DURING STEADY PERIODS (COMPUTER OUTPUT AVERAGES, UNCORRECTED AND UNRECONCILED)

	Dec. 19 19:00-23:00	Dec. 19-20 23:00-03:00	Dec. 20 03:00-07:00	Dec. 20 07:00-10:00
Fuel oil rate kg/hr (lb/hr)	3,300 (7,280)	3,290 (7,260)	3,270 (7,200)	3,280 (7,240)
Electric output, MW	9.7	9.5	9.5	9.5
Air and flue gas to plenum, kg/hr (lb/hr) ^a	12,300 (27,200)	12,300 (27,200)	12,300 (27,100)	12,100 (26,700)
Frothing Air (estimated) kg/hr (kg/hr)	1,360 (3,000)	1,360 (3,000)	1,360 (3,000)	1,360 (3,000)
Limestone feed, kg/hr (lb/hr)	426 (940) ^b	585 (1,290)	576 (1,270)	318 (700) ^c
Gasifier bed temperature, °C (°F)	893 (1,640)	893 (1,640)	904 (1,660)	904 (1,660)
Average pit temperature, °C (°F)	793 (1,460)	788 (1,450)	799 (1,470)	799 (1,470)
Stack gas O ₂ vol %	3.6	3.6	3.5	3.6
Stack gas CO ₂ vol %	11.5	11.6	11.6	11.5
Stack gas SO ₂ vol ppm	29	30	40	53
Stack gas NO _x vol ppm	87	90	92	95
Stack gas opacity, %	33	32	27	20

^aAir/flue gas split not recorded at this stage.

^b585 kg/hr (1,290 lb/hr) while feeding steadily but tripped out at 19:00 to 20:10 and on restart fed at 490 kg/hr (1,080 lb/hr) from 20:10 to 21:15 while being checked out.

^c581 kg/hr (1,280 lb/hr) while feeding steadily, however the feed system tripped out and was shut down at 8:50 until 10:50.

TABLE B-5. ANALYSES OF GASIFIER BED SAMPLES

Sample No.	Day	Hours:min	Mode	Sulfur %	Carbon %
1	18	-	Combustion	1.49	ND
2	19	15:10	Gasification	4.56	1.40
3	19	16:40	Gasification	4.98	0.62
4	19	17:15	Gasification	5.39	1.70
5	20	12:10	Gasification	9.28	2.16
6	20	16:35	Gasification	9.12	ND
7	20	17:10	Gasification	8.63	0.24
8	20	17:10	Gasification	8.68	ND
9	21	8:00	Gasification	8.36	ND
10	21	8:00	Gasification	8.36	ND

Notes: Samples were taken from bottom of front except for sample 9 which was taken from mid-top.

TABLE B-6. ANALYSES OF CYCLONE FINES
(14:00 HR ON DECEMBER 20)

	CY-101A	CY-101B
Sulfur, wt %	5.36	3.08
Carbon	4.68	7.61
More than 20 mesh (74 μ)	29.6	ND
325-200 Mesh (45-74 μ)	16.4	ND
Less than 325 mesh (45 μ)	Trace	--

Surprisingly, the sulfur content of cyclone fines was much lower than that of the gasifier bed material and also there were no fines in the pan from the sieve analysis indicating that the considerable air leak up the cyclone/dip legs was causing some regeneration and also elutriation of fines.

Inspection

A complete inspection of the gasifier/regenerator and boiler was conducted to determine the effects of the gasifier run.

Results of this inspection revealed the following:

Gasifier--

- Coking in several pits
- Glazing of refractory
- Damage to brick work at gasifier to regenerator transfer slots
- Damage to brick wall separating transfer slots from regenerator to gasifier
- Two bed thermocouples bent near the base
- Accumulation of bed material at entrance to product gas conduits
- Accumulation of fines on wall of cyclone CY-101A.

Coking occurred in five pits, Nos. 5, 13, 14 and 15 on the front side of the gasifier and in two pits, Nos. 10 and 15 on the rear side. The pits were not badly coked with the exception of No. 15 where a large quantity of coke

and bed material accumulated into a lump, filling the pit and extending vertically up the end wall which separates the gasifier and regenerator to a height of approximately 0.6 m (2 ft). All pits were cleaned.

The glazing on the gasifier wall proved to be a thin layer of an unknown material deposited on the walls. This material flaked off quite easily and did not appear to be damaging to the gasifier walls in any way.

The only significant refractory damage in the gasifier was the loosening of some brick work at the top of the wall section containing the transfer slots, and the separation and fracture of the short wall between the outlets of the transfer slots from regenerator to gasifier. Both of these failures were repaired.

Two bed thermocouple wells became bent near the base through an angle of 10 to 20°. Both were repaired by applying heat at the bend and straightening. All dust and bed material were removed.

Boiler--

The boiler was last cleaned prior to the 66 hour combustion run. Therefore, at the beginning of the run, there were significant deposits of fines on convection pass tubes in the air heater. At the end of the run there was an accumulation of fines at the bottom of the boiler, and at the base of the convection pass tubes and on the air heater.

The pilot in the No. 3 burner was inoperative.

Boiler Efficiency--

Attempts were made to establish boiler efficiency during combustion and gasification. Unfortunately, the flow rates indicated for superheated steam, feedwater, and natural gas were questionable as discrepancies existed where there should have been none (i.e., with no blow down, feedwater and steam flow should have been equal; however, in some cases steam flow was 10 percent lower than feedwater flow). Therefore, determination of efficiency by the loss method was probably the most accurate. The approximate results are shown in Table B-7 for 3 periods for which sufficient unreconciled data are available.

The last line in Table B-7 refers to data obtained at about 10 MW gross on natural gas only (prior to dust deposition) obtained in March 1979.

Problems

The problem of utmost importance encountered during combustion and gasification was that of dust loading to the cyclones being higher than anticipated. Material was lost from the gasifier bed at a rate of approximately 300 kg/hr (670 lb/hr). As a result, a substantial quantity of dust entered the boiler and accumulated on the convection pass tubes. From available data, it is not possible to determine cyclone efficiency exactly, but a rough approximation indicates about 80 to 85 percent. Therefore, a substantial quantity was retained in the boiler and eventually would have seriously affected boiler performance.

TABLE B-7. BOILER EFFICIENCY RESULTS

Mode	Percent heat supplied		Boiler efficiency, percent	
	Natural gas	Product gas	Loss method	Input-output method
Combustion	85	15	78.4	84.0
Gasification	33	67	84.7	79.8
Gasification	--	100	84.9	83.5
Comparison	100	--	83.0	ND

The dust which accumulated in the boiler and air heater resulted in high flue gas temperatures exiting the air heater (highest was 243°C or 470°F compared to a clean boiler value of 157°C or 315°F from natural gas). Although these high temperatures did not endanger the stack, they did cause high flue gas recycle baghouse temperatures and inhibited use of flue gas as an inerting medium to the gasifier.

An additional difficulty encountered was the repeated jamming of the spent material receiver-filter rotary feeder. Clearances on the feeder were reset, and thermal overloads and overload block were replaced. However, a problem still existed as the feeder continued to trip. The cause of the problem was being investigated.

Another difficulty was encountered during the initial startup of the unit when the flue gas recycle baghouse auger drive motor burned out. This was the third time this had occurred. It was later determined that the thermal overload circuit had been bypassed. This problem was rectified and the motor was removed. The loss of this motor necessitated manual operation of the auger during the test run.

PART 2: FUEL OIL GASIFICATION (RUN 2)

Preparation

The gasifier regenerator and boiler were cleaned out and the damage to the partition wall between the gasifier and the regenerator was repaired. The two bent thermocouple sheaths were straightened since there were no signs of damage.

Startup

For the initial startup, approximately 91.4 cm (36 in) of sulfated bed from Run 1 was preheated with overhead burners and hot flue gas. By the afternoon of January 26, 1980, all systems were checked out and both the gasifier bed and the refractory were preheated to about 200°C (400°F). Early on January 27, there was an attempt to feed lignite into the gasifier but this was frustrated by wet lignite in the rotary feeders and poor fluidization. This was solved by drying and increasing the flue gas flow, respectively, following which 3,384 kg (7,460 lb) of lignite was run into the fluidized gasifier bed. At about 12:15 smooth ignition occurred in the gasifier bed and duct oxygen soon decreased to 0 percent. Admission of air to the plenum until duct oxygen increased to 0.2 to 0.3 percent caused a rapid temperature increase in the gasifier bed, 816°C (1500°F) being reached at 13:08. Air to the plenum was increased further and at 13:18 eight oil guns were put into service at 544 kg/hr (1,200 lb/hr) and a total air and flue gas flow of 15,290 kg/hr (33,700 lb/hr). A large differential between the gasifier and pits of about 122°C (220°F) was traced to lack of frothing air which was put in service at 14:21, while the total air and flue gas flow was decreased to 12,470 kg/hr (17,500 lb/hr). Limestone flow at this stage was about 320 kg/hr (700 lb/hr).

Change to Gasification

Several hours were spent freeing the cyclone dip legs and spent solids disposal lines. By 19:30 all the CAFB ancillaries were operational and the plant was stable at the following conditions:

- Fuel oil rate = 394 kg/hr (866 lb/hr)
- Air and flue gas to plenum = 10,864 kg/hr (23,900 lb/hr)
- Frothing air (estimated) = 1,364 kg/hr (3,000 lb/hr)
- Lignite injector flue gas (estimate) = 2,730 kg/hr (6,000 lb/hr)
- Limestone rate = 487 kg/hr (1,072 lb/hr)
- Gasifier temperature = 857°C (1575°F)
- Pit temperature, average = 732°C (1350°F)
- Gas duct O₂ = 6 volume percent
- Flame in burners No. 1 and 4, pilot only in No. 3
- Oil guns in use: 2, 5, 8, 11 and 12 rear; 2, 5, 8, 12 and 13 front

Over the next 10 minutes, the oil flow was increased steadily with the following results:

<u>Time</u>	<u>Oil flow, kg/hr (lb/hr)</u>	<u>Notes</u>
19:36	855 (1,880)	Duct O ₂ 4.1 percent and decreasing
19:38	1,855 (4,080)	Duct O ₂ reading zero
19:40	2,440 (5,370)	Have ignition on No. 3 burner
19:42	-	Have ignition on No. 4, N.G. valve closed
19:45	3,730 (8,220)	On steady gasification
20:00	-	No. 1 (N.G.) burner taken out of service
20:17	4,370 (9,610)	Igniters taken out of service on No. 3 and 4

At this stage, the boiler was running at 13.5 MW gross on product gas only with 24 oil guns in service.

Maximum Continuous Rating (MCR) Run

Over the next hour, the oil flow was gradually increased and flue gas to plenum was gradually backed out: all the recycle gas to plenum was backed out by 20:48 and the oil rate reached the maximum continuous rate of 273 kg/hr (600 lb/hr) per gun, with 28 guns in service, at 21:22 on January 27, 1980.

Oil flow of between 7,257 and 7,670 kg/hr (16,000 and 16,910 lb/hr) at a plenum air flow of 13,200 kg/hr (29,100 lb/hr) and 21 to 22 MW gross generation were maintained for 1.5 hours. Limestone flow was increased to 370 kg/hr (1,700 lb/hr) and maintained at that rate until sulfation. Conditions remained stable with no plant or operational problems.

At 23:22 the gasifier plenum air rate was reduced to 11,545 kg/hr (25,400 lb/hr) and a minute later, as the gasifier temperature started dropping, the oil rate was reduced. Throughout the MCR run there were no problems with any of the CAFB or boiler equipment.

Load Following (10 MW, 23:20 on January 27 to 22:30 on January 29, 1980)

Over the next 25 minutes, the oil rate was reduced to 2,780 kg/hr (6,120 lb/hr) at 10.5 MW at which point (23:48) the load increased to 12 MW and the oil rate had to be increased to stem the fall in the superheater temperature. This change set the pattern for the next 2 days during which the generators were sending out about 10 MW (9 to 12 MW) and the oil rate varied in the range of 2,770 to 3,400 kg/hr (6,100 to 7,500 lb/hr), averaging about 3,270 kg/hr (7,200 lb/hr). The gasifier temperature remained mostly in the range of 860°C to 877°C (1580°F to 1610°F) and the number of oil guns in service varied from 12 to 20 but was usually 14. There were quite a few oil gun changes during this period whenever a pit temperature dropped or appeared to drop. In most cases the low temperature pit recovered soon after an oil gun change. In general there appeared to be little difference in temperature between pits in service and pits out of service. There were also a number of small but sudden decreases in oil flow rate which were identified as mostly caused by blockages of the small oil gun orifices by debris. Since the installation of the flexible connectors and oil gun rod-outs, clearing either the orifice or any gun restriction caused no problems.

Base Load Operation at 15 MW (22:30 on January 29 to 6:35 on January 30, 1980)

An operational change was made at 22:30 on January 29 to increase the load on the boiler and by 22:37 the load stabilized at 15 MW with oil flow at 4900 kg/hr (10,800 lb/hr). This condition was maintained until 6:35 on January 30, when the gasifier tripped out on a low bed velocity alarm with a consequent loss of flame and shutdown. The trip was a direct consequence of a problem that manifested itself somewhat earlier; at 3:20 the average rear bed temperature dropped suddenly without any discernible cause. After carrying out some detailed checks of the logged data, it was noticed that the zone 5 thermocouples (see Figure B-1) were all reading well below 370°C (700°F). The indicated bed velocity includes a correlation for the average bed temperature, therefore an indicated (but erroneous) sudden drop in bed temperature from 860°C to 727°C (1580°F to 1340°F) caused an indicated drop of 12 percent in the fluidizing velocity. At 6:30 there was a further sudden drop in the indicated gasifier bed temperature to 593°C (1100°F) as zone 2 thermocouples went out of action. The effect on the indicated fluidizing velocity was to drop it 25 percent below the actual velocity but, what is more important, to reduce the indicated bed velocity below the low trip level of 0.61 m/sec (2 ft/sec). An attempt was made to switch over to natural gas but was thwarted by a statutory 5-minute purge interlock: it was not necessary in this case but was executed automatically by the interlocks as soon as the attempt was made to light the natural gas burners. A further attempt to light the natural gas burners at the end of the 5-minute purge also failed because the control logic is such that a low gasifier bed temperature causes product gas burner flame detector shutters to close (to prevent them from detecting the flame from natural gas burners). As product gas was the only fuel being utilized, the boiler, and subsequently the gasifier, tripped on loss of boiler flame, which was caused initially by the fuel oil trip which was caused by the low bed velocity.

The enforced shutdown was utilized to carry out some repairs on the flue gas baghouse auger, and to physically disconnect the failed thermocouples. The minicomputer program logic was altered to cut out the failed thermocouples from the averaging algorithm.

Restart, More Gasification of No. 6 Fuel Oil, and Sulfation

The boiler was put back into service at 10:20 on January 30, with natural gas through Nos. 2 and 3 burners, igniter in No. 4 burner and generating about 7 MW. Oil injection commenced through six guns at 11:20 with a gasifier bed temperature of 829°C (1525°F). Product gas ignition occurred at the No. 4 burner at 11:22 with an 1,810 kg/hr (4,000 lb/hr) oil rate and 9,530 kg/hr (21,000 lb/hr) of air only to the plenum. A switch over from the natural gas to product gas proceeded smoothly and rapidly; by 11:35 the boiler was at 10 MW gross on product gas only at an oil rate of 3,356 kg/hr (7,400 lb/hr) and an air rate of 11,340 kg/hr (25,000 lb/hr). During the changeover, the gasifier bed temperature rose rapidly to a peak of 899°C (1650°F) at 11:30 and then settled at 885°C (1625°F). Between 12:20 and 12:50, the oil rate was ramped up to 4,540 kg/hr (10,000 lb/hr) with 15 MW out of the boiler. It was planned for

these conditions to be maintained for a considerable period, but at 23:15 zone 1 thermocouples failed and at 2:00 on January 31, zone 6 thermocouples failed. In each case, the faulty thermocouples were removed rapidly from the averaging algorithm so that no shutdown occurred, but after 2:00 there were only two thermocouple stacks left in service so that the next failure would have tripped the boiler out unless some remedial steps were taken. Consequently, the No. 1 burner was put in service on natural gas, the oil flow was reduced to 3,630 kg/hr (8,000 lb/hr) and preparations were made to gradually transfer from product gas to natural gas and to sulfation. The boiler load at this point was swinging quite a lot; therefore, the ramping down was delayed until the boiler swings calmed down before proceeding with further changes. At 5:00, with the boiler operation stable, oil flow was reduced to 2,860 kg/hr (6,300 lb/hr), and 11,800 kg/hr (26,000 lb/hr) of air and flue gas to the plenum. The transfer of load proceeded in small steps until 6:37, with the oil rate at 1,900 kg/hr (4,200 lb/hr) and half the boiler load provided by natural gas, the CAFB was ready for a changeover to sulfation. The gasifier temperature was 849°C (1560°F) and air and flue gas (mostly flue gas) flow was 10,610 kg/hr (23,400 lb/hr).

At 6:34 oil flow was reduced sharply causing a loss of product gas flame by 6:36 when the oil flow dropped to 360 kg/hr (800 lb/hr). At 6:38 oil flow was reduced to zero, the remaining oil guns taken out of service and sulfation was started.

Gasifier temperature immediately started rising:

<u>Time</u>	<u>Notes</u>
6:45	Closed gasifier air insulation damper, gasifier temperature 954°C (1750°F)
6:51	Peak temperature of 1024°C (1875°F) reached, have a mild regeneration
6:53	Blowing nitrogen into plenum
7:04	Gasifier temperature now 968°C (1775°F) and dropping
7:12	Backed off flue gas recycle gas (recycle O ₂ high)
7:15	Slumped bed at 0.52 m/sec (1.7 ft/sec)
7:23	Fluidized on flue gas; temperature rising again
7:26	Slumped; temperature 968°C to 982°C (1775°F to 1800°F)
7:31	Removed nitrogen purge from plenum
7:41	Closed off frothing air; shut down pulsers
7:59	Recycle fan taken out of service

At 12:50 on January 31 (after a shutdown of just under 5 hours) the gasifier bed was refluidized with flue gas only; the gasifier bed temperature at this point was 899°C (1650°F) and began increasing because the flue gas contained too much oxygen. Boiler oxygen was reduced and by 13:30 at a boiler oxygen of 2.5 percent and a recycle flow of 8,120 kg/hr (17,900 lb/hr) the gasifier bed temperature was remaining steady at 885°C (1625°F) and the stack SO₂ at 200 ppm.

Sulfation was complete and burn forward of product gas conduits finished by 15:00. At 15:01 two oil guns, 5 rear and 7 front, were put into service at 272 kg/hr (600 lb/hr) and gasifier bed temperatures were 680°C (1250°F) and falling rapidly. At 15:09, with 5 rear and 7 front pits at 554°C (1030°F), the oil guns were taken out of service. The two guns were put back into service, at a low rate, at 15:14 and taken out again when the gasifier was shut down at 15:35.

Regenerator

Several attempts were made throughout the run to put the regenerator into service. However, these attempts were fruitless and were impeded by the inability to transfer material to the regenerator. At one point, regenerator temperatures were sufficiently high to enable oil to be injected and combusted. Combustion was maintained for a short period of time and then terminated, as it was believed that sufficient material transfer was being maintained to continue combustion. However, this assumption was incorrect and regenerator temperatures were lost. Repeated efforts to increase material transfer by rodding out transfer slots and using high-pressure nitrogen to assist flue gas in transferring material met with limited success. Gasification was terminated before additional efforts to start the regenerator were possible.

Emission Data

During gasification, the gasifier was operated at three load conditions: 9 to 11, 14 to 15, and 20 to 22 MW. During each of these load conditions, maintenance of stable boiler conditions was attempted but there were some variations made in order to maintain superheat temperatures. Therefore, emission data, which were recorded approximately every half hour, were grouped into nine sets, each corresponding to a boiler load or transition from one load to another. The conditions during each of these sets are tabulated in Table B-8.

The gaseous emission data for each test period are shown in Table B-9. This table indicates the highest and lowest concentrations observed and the mean value for the following gases: carbon dioxide, carbon monoxide, oxides of nitrogen, sulfur dioxide, and oxygen. In addition, the average mass loadings for NO_x and SO₂ are shown.

When reviewing the data, it is readily apparent that SO₂ emissions were somewhat higher than observed during the first gasification run in December 1979. However, the initial gasifier bed material utilized for this run was the same as that utilized for Run 1. Also, the sulfur content of the bed at the end of Run 1 was in excess of 8 percent of the sulfated bed. Further, towards the end of this run, some bed samples were determined to have in excess of 20 percent sulfur. This is not surprising because the regenerator was not in operation during this run, and apparently the gasifier bed was nearing a saturation level for the quantity of reactive material in the bed (theoretical sulfur content of a totally saturated bed would be 44 percent).

TABLE B-8. GASIFIER AND BOILER CONDITIONS DURING RUN 2

Period	Date	Time	Hour of gasification	Mode	Boiler load (MW)	Stack O ₂ (vol. %)	Oil flow, kg/hr (lb/hr)
0	27	20:00 ^a	0	Startup	10-11	NR ^b	3,630 (8,000)
1	27	20:00-21:30	0-1.5	Load change	10-20.5	7.6	3,630-7,670 (8,000-16,900)
2	27	21:30-23:30	1.5-3.5	MCR ^c	20.5-22	5.5	7,670 (16,900)
3	27	23:30-24:00	3.5-4.0	Load change	22-10	5.5-7.4	7,670-3,000 (16,900-6,600)
4	28-29	00:00-22:00	4.0-50.0	Steady	10	5.8-7.8	3,340 (7,350)
5	29-30	22:00-6:30	50.0-58.5	Steady	14-15	4.0	4,880 (10,750)
-	30	6:30-11:30	-	Shut down	0	NR	0 (0)
6	30	11:30-13:00	58.5-60.0	Restart	0-15	NR	0-4,770 (0-10,500)
7	30-31	13:00-2:30	60.0-73.5	Steady	15	5.4	4,580 (10,080)
8	31	2:30-4:30	73.5-75.5	Load changed	15-11	NR	4,540-3,630 (10,000-8,000)
9	31	4:30-6:30	75.5-77.5	Load transfer	11	NR	3,630-1,815 (8,000-4,000)
-	31	6:30	77.5	Start sulfation	NR	9-2.5	0 (0)

^aArbitrarily when all natural gas was taken out of service.^bNR = Not recorded.^cMCR = Maximum continuous rate.^dNatural gas lit at burner No. 1

TABLE B-9. GASEOUS EMISSIONS DATA, RUN 2

Test period	Hours of gasification	Oxides of Nitrogen		Sulfur Dioxide		Carbon monoxide concentration (ppmv)	Carbon dioxide concentration (Vol %)	Oxygen concentration (Vol %)
		Concentration (ppmv)	ng/J	1b/10 ⁶ Btu	Concentration (ppmv)	ng/J	1b/10 ⁶ Btu	
1	0 - 1.5	113	84.8	0.197	302	315	0.733	8.97
2	1.5 - 3.5	131	86.1	0.200	498	456	1.06	7.4
3	3.5 - 4	82	73.6	0.171	290	363	0.843	11.0
4	4 - 50	77	61.5	0.143	155	171	0.397	9.64
5	50 - 58.5	100	60.7	0.141	311	262	0.609	6.19
6	64 - 65	106	68.4	0.159	250	225	0.522	7.1
7	65 - 78.5	104	68.0	0.158	281	256	0.595	7.27
8	78.5 - 80.5	107	72.7	0.169	188	169	0.392	7.10
9	80.5 - 82.2	104	77.0	0.179	133	59.8	0.139	8.9

Figure B-2 shows the sulfur removal efficiency over the length of the run. The highest levels of SO_2 were observed during the initial 4 hours of operation, at which time the gasifier was in a transition from 10 MW on natural gas to 10 MW entirely on product gas, then from 10 to 22 MW stabilized at 20.5 to 22 MW, and reduced to 10 MW. During the ensuing 46 hours, SO_2 emissions dropped significantly while gasifier loading was at 10 MW. When the gasifier load necessitated an increase in fluidizing velocity and fuel flow, the total sulfur input increased and the residence time decreased. Therefore, as there was initially substantial sulfur on the bed material, it was not unexpected that the sulfur capture suffered. Throughout gasification, the sulfur removal efficiency varied from 50 percent (during maximum continuous rating [MCR]) to 93 percent (during extended operation at 10 MW). Throughout the operation, fresh limestone was fed to the bed at a rate of 773 kg/hr (1,700 lb/hr) (1.25 to 2.8 stoichiometric calcium to sulfur) and there was a considerable carryover of fines into the boiler.

Throughout gasification, oxides of nitrogen emissions from the boiler remained low. For higher gasifier loads, NO_x emissions were slightly higher than for reduced loads. This observation has been documented many times. The highest NO_x emission observed was during MCR tests (21 to 22 MW) and reached a value of 143 ppm or 108 ng/J (0.251 lb/10⁶ Btu). The NO_x data for both gasification runs are plotted in Figure B-3 along with emissions from the natural gas tests.

Emissions of particulate matter were monitored continuously as opacity and on a mass rate basis in one particulate test (EPA Method 5). The opacity measurements are shown in Figure B-4. It is apparent from the data in this figure that during a transient condition (change in load or fluidizing velocity, startup, or shutdown), opacity increased significantly. During stable operation, however, opacity levels stabilized and remained at or below 20 percent while the gasifier maintained a load of between 10 and 15 MW. During operation at 21 to 22 MW, opacity was high (50 to 75 percent). However, it was dropping rapidly and had not leveled out prior to reducing load. The results of the Method 5 particulate test indicated the mass loading of 70.6 ng/J (0.164 lb/10⁶ Btu). The opacity during this test was constant at 15 percent.

Analysis of Gasifier Bed Material

Throughout combustion of coal and oil, and gasification of oil, a total of 106 gasifier bed samples were taken from two bed sample ports every 1 or 2 hours. Of these, 46 were analyzed for carbon, hydrogen, and sulfur and six of the samples were sieved to determine size distribution. A plot of carbon on bed material vs. time is shown in Figure B-5. Notations describing operating conditions are also indicated. The method of analysis used to determine carbon and hydrogen is that specified by ASTM modified by including a warmup period with a nitrogen purge to release any CO_2 from recarbonization and water from hydration of the sample during sample preparation (grinding, sieving, etc.).

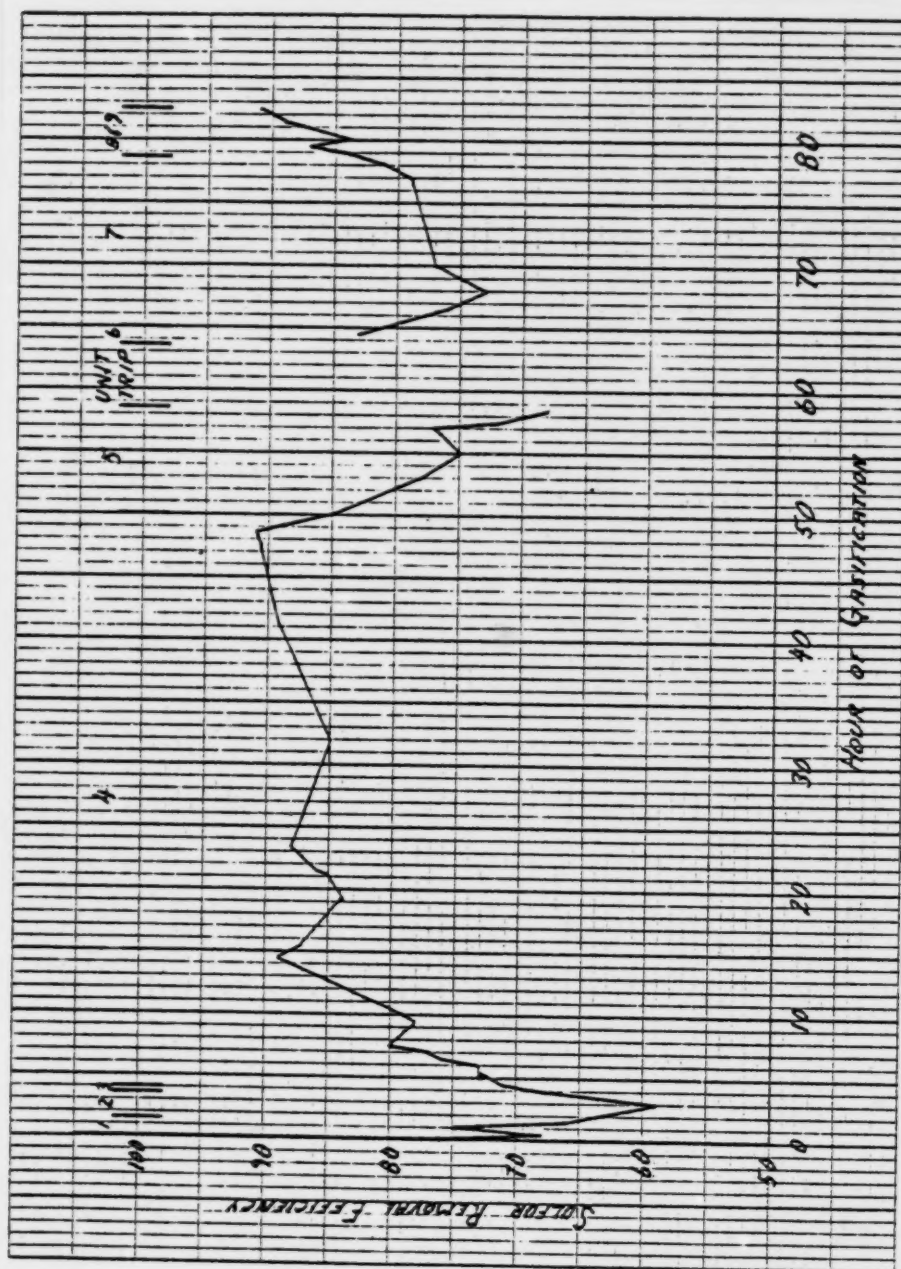
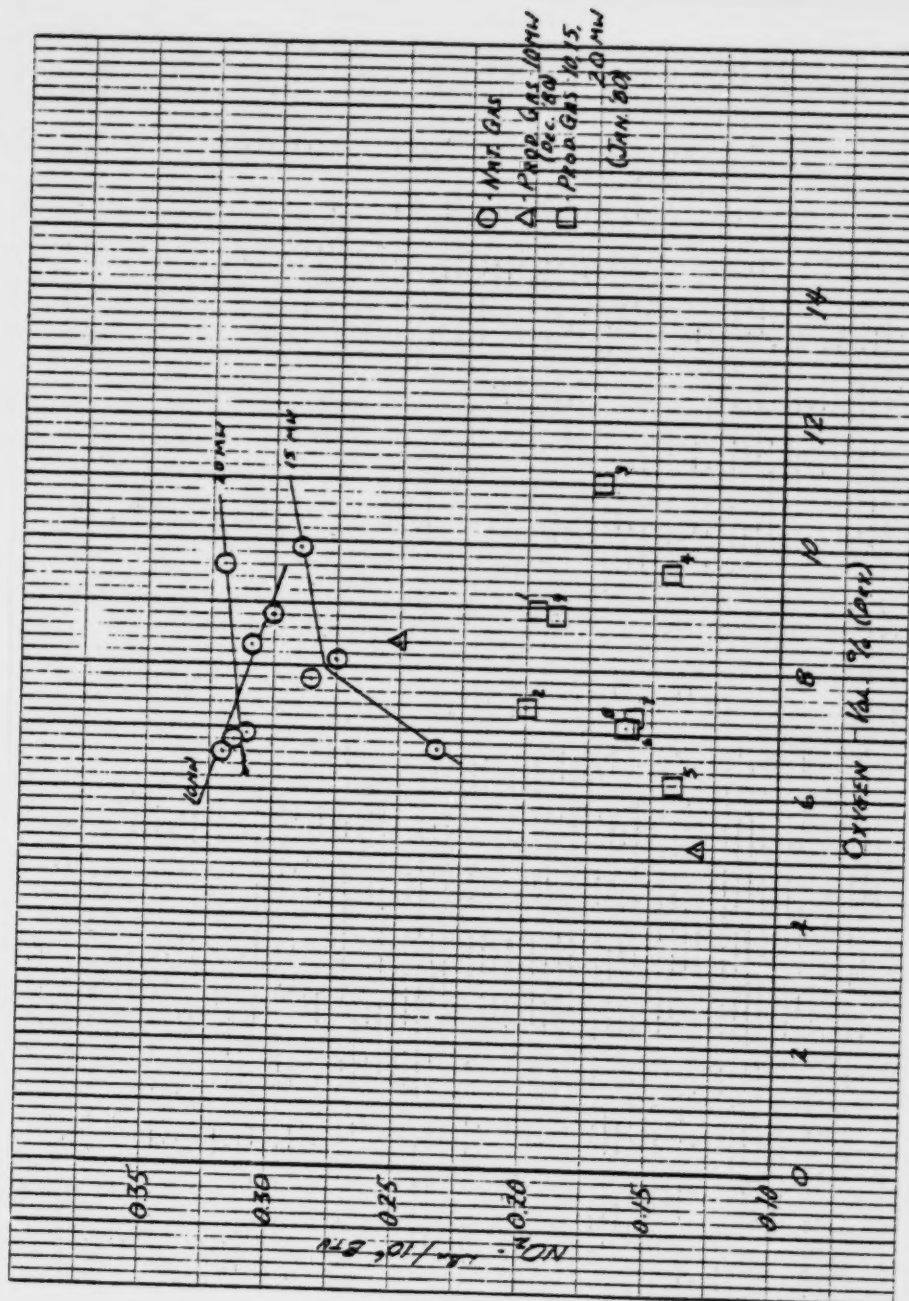


Figure B-2. Sulfur removal efficiency vs. time, Run 2.



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Figure B-3. NO_x emissions during Run 2.

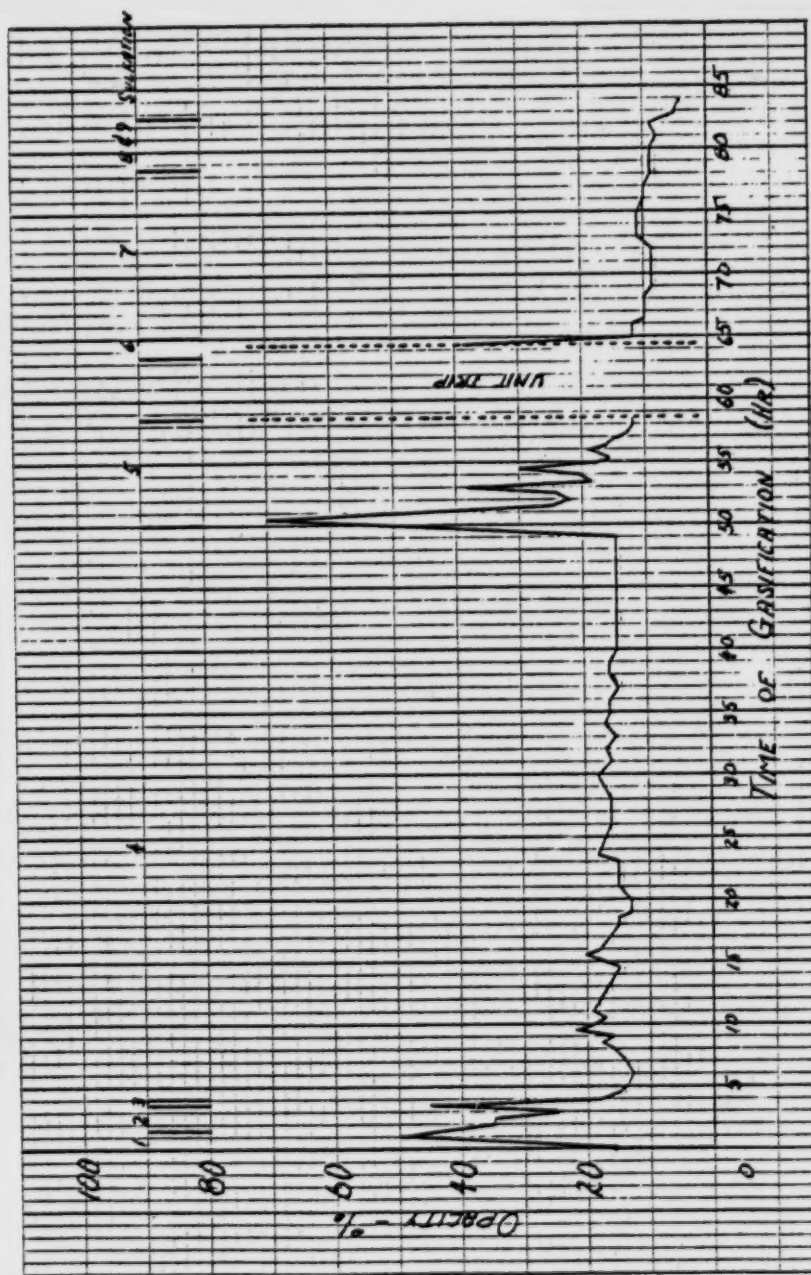
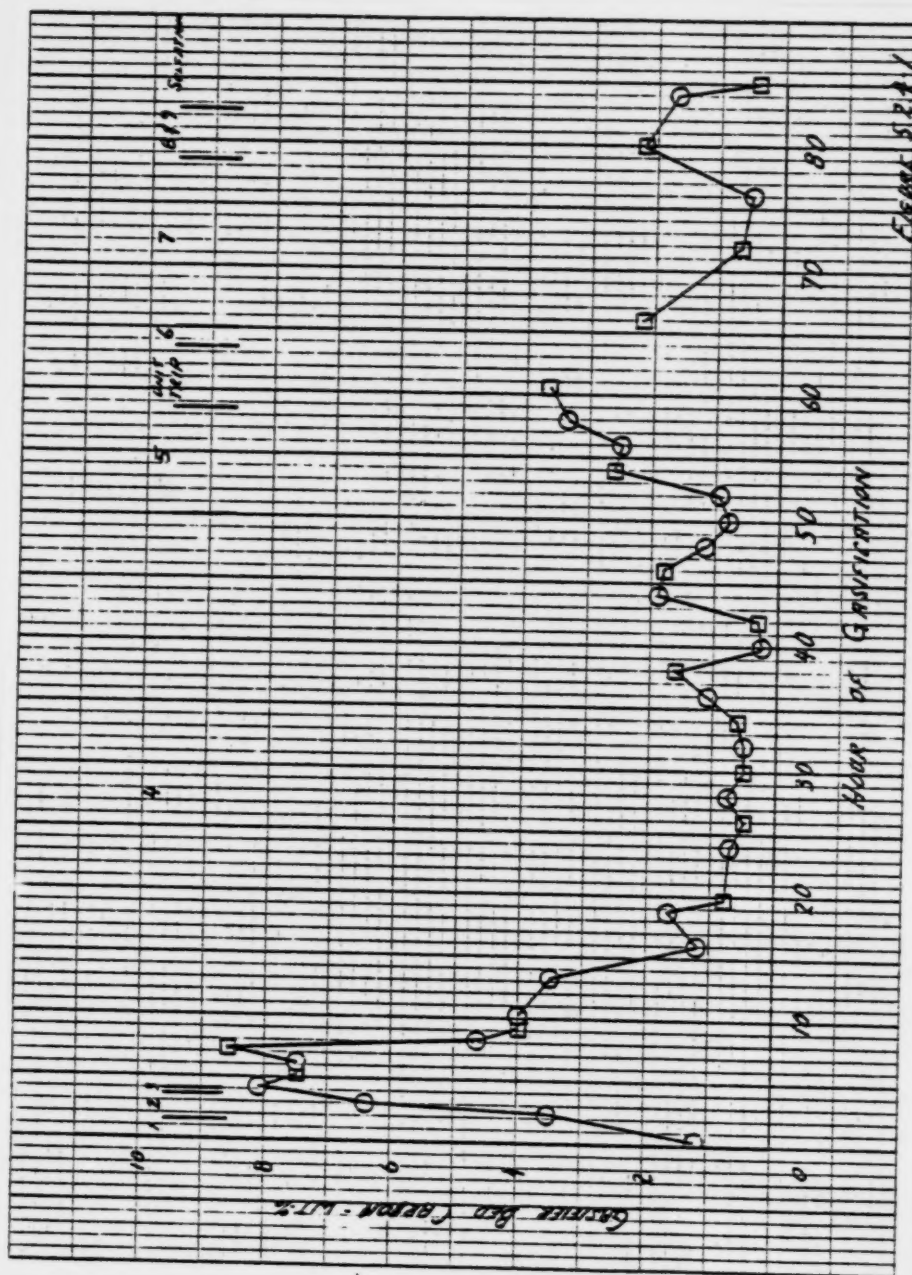


Figure B-4. Opacity during gasification, Run 2.



The data shown in Figure B-5 of gasifier bed carbon content vs. time indicate a rapid accumulation of carbon in the bed after initial startup. The carbon inventory remained very high during the initial 7 hours of gasification, then dropped by approximately 50 percent. After the thirteenth hour of gasification, bed carbon again dropped by approximately 50 percent. According to ERCA, during oil gasification, bed carbon levels should typically be in the 2-percent range for proper gasifier stoichiometries (approximately 20 percent). However, throughout this run, gasifier stoichiometries were initially considerably lower than optimal, possibly as low as 16 percent. Therefore, it is not unusual that bed carbon levels were higher than normal.

The hydrogen content of the gasifier bed material samples fell within a range of 0.07 to 1.2 percent. The trend of bed hydrogen generally followed that of the carbon; that is, the higher hydrogen concentrations were in samples containing high carbon.

Results of the sulfur analysis of bed material are shown in Figure B-6. Twenty-one results are shown on this plot. The procedure recommended by the suppliers of the equipment needed to be modified, by reducing the sample size, to prevent overloading the detection cell; the results were therefore less accurate at the high sulfur level. Also, since the determinations involved much trial and error and several redeterminations, not all the available samples were analyzed. Sieve analyses of several gasifier bed samples did not indicate any trends with time; the average distribution of six samples is shown in Table B-10.

TABLE B-10. SIEVE ANALYSIS OF GASIFIER BED MATERIAL (MEAN OF SIX)

Sieve No.	Particle size (μm)	Retained on screen (wt %)	
		Mean	Standard deviation
8	2380	0.7	1.7
10	2000	7.2	1.2
12	1680	7.4	0.9
14	1410	12.5	1.3
16	1190	11.4	1.8
20	841	29.0	2.9
30	595	20.9	2.3
50	297	10.7	2.8
Pan	-	0.2	0.05

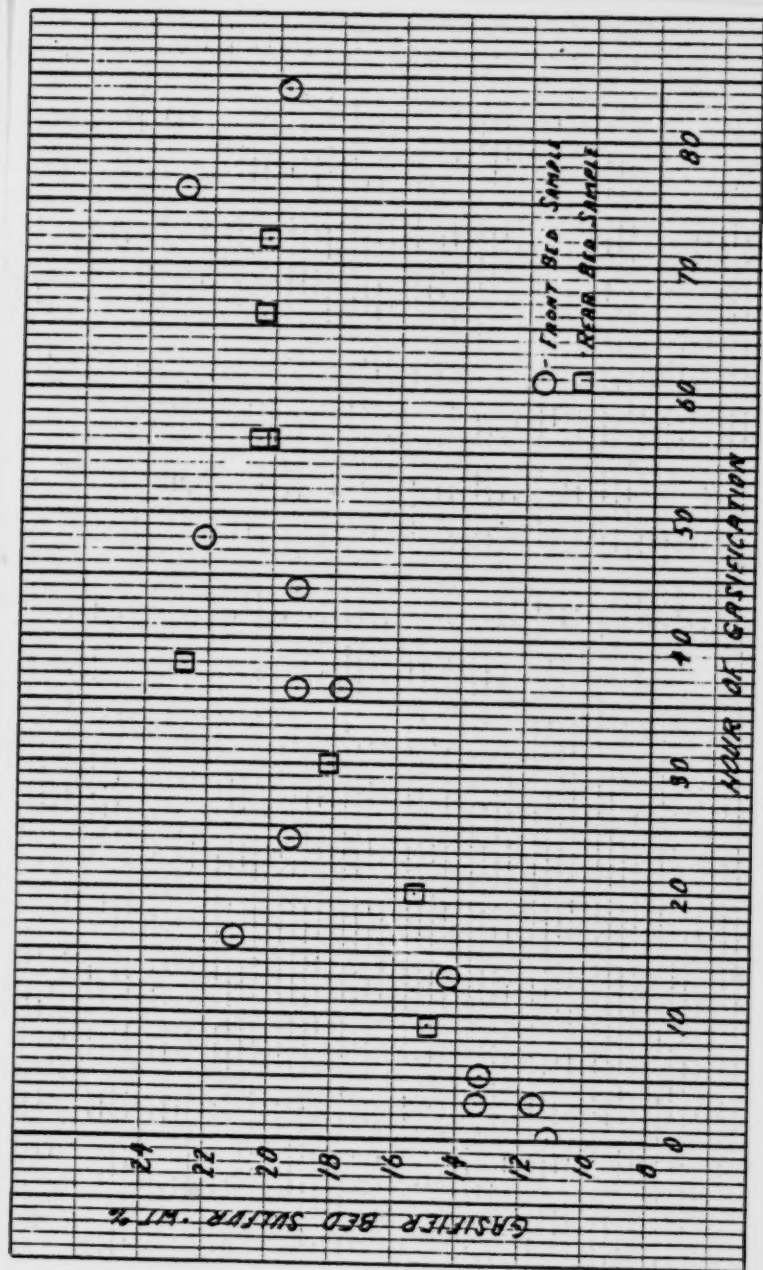


Figure B-6. Gasifier bed sulfur vs. gasification time, Run 2.

When compared to Tables A-2 and A-4, it is seen that the bed material is of a smaller mean size than fresh limestone, but essentially has the same quantity of fine material (that smaller than 30 mesh). This is the result of larger limestone particles fracturing during calcination and of fine material being blown out of the bed.

Boiler Operation

Throughout gasification, boiler operation was basically stable. However, the air heater exit (flue gas) temperature increased approximately 38°C (100°F) and some difficulty was encountered in maintaining final superheat temperature. Both of these difficulties were the result of particulate deposits on boiler and superheat tubes, while the latter problem was apparently complicated by a leaking steam attenuator valve (scheduled to be repaired).

Boiler efficiency during product gas combustion was actually slightly higher than during natural gas combustion despite the duct carryover. Table B-11 shows boiler conditions and efficiencies before and during gasification. Efficiency increased while burning product gas because of higher energy loss from moisture formed during combustion of natural gas. It was apparent that efficiency during product gas combustion decreased with time because of increasing flue gas temperatures and increased tube fouling.

Problems With Gasifier/Regenerator

The main problems encountered during operation were related to gasifier stoichiometry, material transfer between gasifier and regenerator, and thermocouple failures. The low level of stoichiometry in the gasifier (sometimes as low as 16 percent) was indicated by low temperatures in specific pits during gasification caused by coke covering the pit thermocouples. This was not encountered during later operation.

The inability to transfer material between the gasifier and regenerator was apparently the result of high oxygen and moisture concentrations in the transfer gas (flue gas). This caused the material to react within the slots and ultimately to agglomerate.

The thermocouple failures, which ultimately caused termination of the run, came in two stages. In the 55th hour of gasification, the zone 5 bed thermocouple, consisting of three duplex couples in a 76.0-cm (30-in) well, failed followed shortly by the zone 3 thermocouple (see Figure B-1 for location). At this time, the unit tripped on loss of flame in the boiler caused by a false low bed velocity alarm. After restart, the gasifier was operated for 14 more hours, at which time the zone 1 thermocouple failed followed by the zone 6 thermocouple. As these failures left but two thermocouples in service, the run was terminated.

Additional problems which were encountered include coal needle pluggage, other coal system pluggage, refractory cracking in the gasifier, and back leakage through the cyclone rotary valves.

TABLE B-11. BOILER PERFORMANCE--BEFORE AND DURING GASIFICATION

Fuel	Date	Time	Load (MW)	Flue gas O ₂ (%)	Stack temperature, °C (°F)	Efficiency loss method (%)
Natural gas	25	12:10	9.8	7.5	141 (285)	84.25
	25	22:30	9.8	7.0	146 (295)	84.87
	26	04:30	9.8	6.7	203 (398)	81.57
	27	10:45	9.7	6.8	183 (362)	82.61
	27	16:10	11.25	4.5	179 (355)	83.62
	27	18:00	11.25	4.4	177 (350)	83.69
Product gas ^a	27	0-1	11.0	7.6	176 (349)	86.25
	27	1-3	21.25	5.5	193 (380)	86.47
	28	4-18.5	9.8	7.8	182 (360)	85.84
	28-29	21-45	9.8	5.8	182 (360)	86.44
	29-30	51-57	15.0	4.0	207 (404)	86.48
	30	67-76	15.0	5.4	199 (391)	86.12

^aTimes listed for operation on product gas are for hours into gasification.

Coal needle pluggage was a problem from the initial startup. This was a chronic problem because of condensation of moisture from the flue gas which was used as the coal conveyor seal gas. This moisture dampened incoming coal and any bed material pushed up the coal needles, causing the coal to agglomerate and the bed material to react and harden. Moisture from coal conveyor seal gas also condensed on the rotary valves feeding coal from the weigh tables to the coal conveyors. This caused agglomeration of coal fines in the rotary buckets and backing up of coal into the weigh feeders. In one instance, this condition was inadvertently allowed to proceed too long, causing a weigh feeder belt to rip, disabling one-fourth of the total coal feed capabilities.

Back leakage through the gasifier cyclone rotary valves was expected but in practice it rather exceeded the expectations. The leakage was particularly severe through the "A" cyclone valve and caused combustion of carbonaceous material in the cyclone fines with the oxygen in the flue gas used as the transport gas below the rotaries. Several times the dip leg temperature exceeded the cyclone temperature. Pressure tappings were unreliable and blocked several times.

Inspection: Gasifier

Upon inspection of the gasifier following the run, it was observed that all four thermocouples had failed, at or near the floor of the gasifier. The thermowells were basically intact above the break point and were found lying horizontally on the gasifier floor. Samples of the thermowells were sent to the FWEC metallurgical laboratory for analysis. The consensus of opinion was that failure was caused by fatigue and creep accelerated by high temperatures.

All four transfer slots were found to be filled with a solid mass of bed material. In addition, accumulation of hardened bed material approximately 25.4 cm (10 in) thick was found on top of the entrance to the gasifier-to-regenerator transfer slots. A similar accumulation was found on top of the step wall at the exit of the regenerator-to-gasifier transfer slots. This material was somewhat brittle and was not too difficult to remove. Several pits were partially or completely filled with coke. This caused pit thermocouples to be covered and to indicate low temperatures. Several pits contained "cow pats"--brownish brittle agglomerates about 7.5 to 10.0 cm (3 to 4 in) in diameter and 2.5 cm (1 in) thick. They could have been formed at the end of the coal needles when flue gas containing oxygen was blown through them into a slumped gasifier bed. Two of these "pats" were much larger and blacker and could have been caused either by injection of cold oil or injection of hot oil into a bed that was insufficiently hot.

The refractory damage in the gasifier consisted of some cracking and a small amount of erosion. The major cracking was located in the wall containing the transfer slots. A section of this wall containing gasifier-to-regenerator slots had separated slightly and was protruding approximately 0.635 cm (0.25 in) from the rest of the wall. There was a small hole in the gasifier/regenerator wall, in the center about 2.1 m (7 ft) up. Additional

cracks were found in the gasifier center separation wall. There was also a crack perhaps 2.5 cm (1 in) wide in the front startup burner throat at the interface of gasifier and burner refractories.

Inspection: Boiler

There were considerable dust accumulations on furnace waterwalls, convection pass tubes, and air heater tubes, deposits on the burner throat, and erosion of the refractory brick floor. The deposits of material on the burner throat and furnace and convection pass tubes were similar in nature to those found after the previous run. The material on the burner throat was crusty on the surface and loose beneath the surface. It was easily removed and came off in fairly large chunks. The accumulations on the furnace waterwalls were fairly thin and crusty in most cases. It appeared that most of the material which had accumulated on the waterwalls had fallen off during the run as there was a considerable amount of material on the furnace floor. The material accumulated on the leading surface of the convection pass tube was slightly crusty and very fragile, while material on the rear surface of the tubes was dusty. Accumulations on the front surfaces were up to 0.32 cm (0.125 in) thick, while those on the rear surfaces were very thin. The material on the air heater tubes was a very fine dust, perhaps 0.08 cm (0.03 in) thick.

The damage to the furnace floor was primarily in front of the No. 4 burner. This section of the refractory brick floor had erosion grooves as much as 3.8 cm (1.5 in) deep (the thickness of the original brick was 7.6 cm (3 in)). The damage was most severe approximately 3.0 to 4.6 m (10 to 15 ft) in front of the burner, tapering off closer to and further from the burner. Approximately 20 percent of the furnace floor was damaged. There was damage in front of the No. 3 burner as material from the gasifier bed had accumulated to a thickness of approximately 5.08 cm (2 in) in front of the burner. A FWEC refractories expert visited the job site, analyzed the damage and prescribed a repair procedure.

There was very little other damage to any other areas of the boiler. Product gas burners appeared to be in excellent condition as did other primary components.

PART 3: FUEL OIL GASIFICATION (RUN 3)

Preparation

Faults found during the inspection after the previous test run were repaired and some minor modifications and checks were made:

- refractory cracks and failures were repaired;
- thermocouple levels were set at:

gasifier: 20.3, 50.8, and 83.8 cm (8, 20 and 33 in)
regenerator: 25.4, 50.8 and 83.8 cm (10, 20 and 33 in);

- pressure taps, similar to those used at ERCA were made and installed. their measured position, above the grid level, were:

gasifier: 6.4 cm (2.5 in) lower, 38.1 cm (15 in) upper

regenerator: 12.1 cm (4.75 in) lower, 36.2 cm (14.25 in) upper;

- several thermocouples, inserted through the refractory wall and protruding only about 5.0 cm (2 in) into the fluid bed, were installed in the gasifier and the regenerator. These could be used as standby in the event of a failure by the main bed thermocouples.
- the burner nose plugs were stroked; and
- some major valve leaks were identified and repaired by welding.

Extensive flow calibrations and checks were carried out in order to attempt to reconcile some apparent flow discrepancies. The main air and flue gas flowmeter was one of the culprits, it was therefore measured with the results as shown in Table B-12.

Following these measurements, water manometers were installed in parallel with most of the flow measuring instruments and a cross-check carried out, both on the air side and the flue gas side.

Early in March 21,950 kg (48,300 lb) of liquid sulfur was delivered, following which the Resox TM plant was checked out, all the control valves were stroked and the sulfur circulation system checked out in preparation for the inclusion of Resox TM in the next test run. Unfortunately, severe vibration in the tail gas fan caused these plans to be postponed.

Plans for Future Modifications

Because of concerns that the existing coal feed system might not function satisfactorily, various alternatives were considered. Testing of an overbed feeding system at the ERCA pilot plant had shown it to be a satisfactory method of feeding lignite into the gasifier. Plans were therefore made to feed coal at the La Palma CAFB plant using the existing rotary valves and weigh belts and connecting via the access hatch in the end wall. Such a system would be a backup in the event of continuing problems with the existing lignite feed system.

Preliminary Startup Attempts

On March 21; the flow tests and repair work having been completed, limestone was loaded into the gasifier. This limestone was somewhat coarser than that used previously and was damp.

March 23; 66.0 cm (26 in) (static) of fresh limestone had been fed in and leveled, the overhead burners had been lit to start warming up the bed.

TABLE B-12. MEASUREMENTS ON THE AIR AND FLUE GAS VENTURI

Direction of measurement	Position	Measurement, cm (in)	
		Throat	Exit
Vertical	7.6 cm (3 in) from rear side	15.1 (5.95)	48.0 (18.88)
	22.9 cm (9 in) from rear side	15.2 (5.98)	47.0 (18.50)
	Center	15.2 (5.97)	47.1 (18.56)
	22.9 cm (9 in) from front side	15.2 (6.00)	47.3 (18.63)
	7.6 cm (3 in) from front side	<u>15.5 (6.09)</u>	<u>48.1 (18.94)</u>
	Mean	15.2 (5.98)	47.5 (18.70)
Horizontal	2.5 cm (1 in) from top	85.4 (33.63)	-
	Center	85.7 (33.75)	86.4 (34.03)
	2.5 cm (1 in) from bottom	85.7 (33.75)	85.6 (33.72)
	1/4 way down	-	85.7 (33.75)
	1/4 way up	<u>-</u>	<u>85.9 (33.81)</u>
	Mean	85.6 (33.71)	85.9 (33.83)
Corner fillets: 4 off, 2.45 cm^2 0.38 in^2) each = 9.81 cm^2 (1.52 in^2) total.			
Throat area = 0.129 m^2 (1.389 ft^2) compared with 0.137 m^2 (1.476 ft^2) nominal.			
Exit area = 0.407 m^2 (4.383 ft^2) compared with 0.428 m^2 (4.604 ft^2) nominal.			

March 24; attempts to fluidize were unsuccessful since too much bed had been placed in the gasifier and the main blowers could not cope.

March 25; problems with plugging of coal injectors delayed further attempts to start. The flue gas blower was left on to keep the bed warm but it was also blowing out limestone fines.

March 26; the gasifier was cooled with air, about 5.0 cm (2 in) of the bed was drained out, the remainder was leveled manually and some lignite was added. The gasifier was closed again and a reverse heating method was tried by closing the nose plugs on burners Nos. 3 and 4 and trying to force the combustion product through the gasifier bed and vent via an open hatch in the plenum. The method did not appear to be successful because there was too much leakage and only a very low flow through the bed. After closing off the plenum and backing off nose plugs there was partial fluidization and one portion of the gasifier increased in temperature to 843°C (1550°F). At this stage there was failure of a drive sprocket on the "B" coal feed system so the coal feed system was shut down for repairs.

March 27; having repaired the coal feed system, an attempt was made to fluidize and combust with only partial success because problems were experienced with the rotary valves at the bottom of the cyclone drains; the rotor webs had developed cracks. The plant was shut down and the rotors were removed for repairs.

Problems and unsuccessful attempts continued until April 1, when oil combustion was initiated using nine oil guns and about 35.6 cm (14 in.) of bed. About the same time the data logging computer crashed, the rotary valve on B cyclone tripped out, there was some difficulty in maintaining adequate pit temperatures (possibly because of coarse limestone) and there was considerable overhead burning - possibly because of the shallow bed. In view of the adverse circumstances the plant was shut down.

Start-up and Run

The gasifier was started up on April 7, 1980 using a method similar to that described previously (prelim, run 1 and run 2) with 56.0 cm (22 in) of fresh limestone and lignite as the initial fuel. The maximum flue gas rate and maximum frothing air were used for the initial lift off of the gasifier bed because the limestone used was fairly coarse. Steady oil combustion was achieved by 22:00 and the gasifier bed appeared to be fully calcined by midnight when fresh limestone injection, at the maximum rate of about 1090 kg/hr (2400 lb/hr) was initiated to build up the gasifier bed to 76 cm (30 in). Gasification was initiated by increasing the oil rate from 470 to 1810 kg/hr (1040 to 6200 lb/hr), while fluidizing with flue gas. Ignition occurred at 17:22 on April 8, 1980. Gasification continued for a total of 144 hours during a 148-hour period with but two shut-downs, both accidental:

- the first trip occurred at 12:10 on April 11, and was caused by an accidental switching off of the main oil pump. Gasification was resumed 8 minutes later with no noticeable upset to the system.
- the second trip occurred when regenerator vent to stack valve was accidentally closed, while checking the ResoxTM plant instruments, at about 16:25 on April 11, causing a high regenerator to gasifier pressure differential and a consequent fuel trip. Since the gasifier temperature was rather high at the time (it rose rapidly from about 871°C to 960°C (1600°F to 1760°F) because frothing air remained on for awhile) there was a possible risk of regeneration on restart, resumption of gasification was delayed for 4 hours.

Gasifier Operation

Throughout the operation the gasifier provided from 50 to 100 percent of the heat input to the boiler. During this time, the gasifier was loaded at 7.2 to 15.0 MW. A variety of gasifier and boiler operating conditions existed throughout the operation. Data collected during the run were grouped into 10 sets, each set corresponding to particular operating conditions. Operating conditions of the gasifier during each set are shown in Table B-13. Gasifier stoichiometry throughout the run was between 19 and 22 percent. Extreme difficulty was encountered in manipulating stoichiometry because of the high rate of leakage through the main air control damper when closed.

During operation of the gasifier, it was noted that the overhead pressure increased steadily with time. At the beginning of the run, the overhead pressure was typically 0.75 to 1.5 kPa (3 to 6 in H₂O) and increased to approximately 3.5 to 3.7 kPa (14 to 15 in H₂O) at the end of the run. During shutdown when the gasifier bed was being sulfated, the overhead pressure decreased considerably, to approximately 0.75 kPa (3 in H₂O). This suggested that the cause of higher overhead pressure was carbon laydown. This carbon apparently burned off during combustion while attempting to sulfate the bed. In addition, at the end of the run, a large piece of material was observed partially blocking the No. 4 burner. The presence of this material undoubtedly increased the back pressure on the gasifier. This material was apparently consumed during sulfation of the gasifier bed as there was no evidence of it during the post operation boiler inspection.

Some salient features of the run are summarized below:

- April 9: the regenerator would not fluidize 66.0 cm (26 in) of static bed but after draining out about half the bed, rodding the slots and fluidizing, the temperature rose sharply to 870°C (1600°F). The regenerator bed height increased while there was a corresponding reduction in the gasifier bed. As regeneration continued, the temperature increased towards 1100°C (2010°F) which was a trip point; the regenerator was therefore slumped to avoid overheating. Restart at a later time failed.

TABLE B-13. GASIFIER OPERATING CONDITIONS AND BOILER PERFORMANCE DURING RUN 3

Time period	Gasification hour	Boiler load (MW)	Total heat input to boiler, 10 ¹⁰ J/hr (10 ⁷ Btu/hr)	Heat to boiler provided by product gas (%)	Oil flow, kg/hr (lb/hr)	Fresh limestone, kg/hr (lb/hr)	Gasifier stoichiometry (%)	Boiler efficiency ^a (%)
1	0-36	10.6	14.42 (13.68)	87	3,004 (6,622)	976 (2,151)	22	85.4
2	45-49	11.5	16.78 (15.92)	67	2,693 (5,937)	771 (1,700)	22	84.5
3	49-63	15.0	20.27 (19.23)	99	4,799 (10,579)	729 (1,607)	20	86.7
4	63-66	11.5	16.34 (15.5)	100	3,894 (8,585)	0 (0)	20	85.5
5	68-71	10.5	12.96 (12.3)	100	3,090 (6,813)	0 (0)	20	85.7
-	71-76	-	-	-	0 (0)	0 (0)	S.D.	-
6	76-110	10.2	15.67 (14.87)	71	2,441 (5,382)	440 (969)	22	84.6
7	110-113.5	12.5	18.98 (18.01)	62	2,794 (6,159)	322 (710)	22	84.8
8	113.5-115.5	15.0	22.47 (21.32)	50	2,693 (5,937)	322 (710)	22	82.7
9	115.5-147	12.4	18.40 (17.46)	65	2,846 (6,275)	380 (838)	21	84.0
10	147-148	11.5	16.57 (15.71)	49	1,950 (4,299)	454 (1,000)	19	84.2

^aBy loss method.

- April 10; adjusting nose plugs in burners No. 3 and 4 gasifier temperature and fluidization on automatic control, included 10 to 15 MW load changes on automatic
- April 11; O.K. except that at 3:45 the solids cooler tripped out on overload; fines were reinjected into the fluidized regenerator (not operational at 320°C (600°F)), until cooler restarted at 11:10. At 12:10 and 16:25 experienced the two gasifier trips described previously.
- April 12; found 1 gallon of fuel oil in regenerator plenum. Drained some regenerator bed and fluidized the remainder with the overhead burner in service. Had regenerator ignition (which caused high stack opacity for 15 minutes) but no transfer. Gasifier O.K. and still on automatic.
- April 13; attempted to place regenerator in service but was unsuccessful. Gasifier O.K. at 10 MW on automatic control.
- April 14; solids cooler malfunction at 13:30; spent fines were injected into the regenerator which was still not operational. Solids cooler back in service at 18:10; cause of malfunction was a plugged ball valve.

Sulfation: slumped and oil out at 22:18 all boiler load was transferred to natural gas (burners Nos. 1 and 2) refluidized at 22:20 at 821°C (1510°F) and attempted to reduce boiler oxygen. At 22:26 gasifier bed temperature peaked out at 913°C (1675°F) and SO₂ at 570 ppm with plenum O₂ at 9.4 percent, 9620 kg/hr (21,200 lb/hr) plenum gas flow with 0.59 m/sec (1.94 ft/sec) fluidizing velocity. There were no problems with either the bed temperature control or excessive SO₂ release throughout the sulfation.

At the end of sulfation the carbon content of gasifier bed material looked high on visual inspection - continued on combustion.

- April 16; continued combustion until 6:30 followed with forced cooling at 110°C/hr (200°F/hr) with air.

Boiler Operation

During the gasification run, the boiler was fired with product gas only for 22 hours. During the remaining operation (122 hours), the boiler was simultaneously fired with product gas and natural gas, product gas supplying from 50 to 87 percent of the heat input to the boiler. Included in Table B-13 are averaged data taken during the gasification run and classified according to the different gasifier and boiler operating conditions. It can be seen that the highest boiler efficiencies occurred during combustion of product gas only, and that, in general, the boiler efficiency decreased with time (primarily because of particulate deposition on waterwall and superheat tubes). These characteristics have also been observed during previous runs.

Early in the gasification run, the burner nose plugs were manipulated to improve flame characteristics and to increase the velocity of gas entering the product gas cyclones. To accomplish this, the No. 3 burner nose plug was inserted all the way, which essentially eliminated flow of gas through that portion of the system. This procedure doubled the velocity of gas entering the "B" cyclone. The increased velocity, however, did not appear to affect cyclone efficiency significantly, as the stack opacity was of a similar magnitude to that experienced in previous runs with both burners in service. Moreover, when the No. 3 burner was returned to service, stack opacity remained unchanged. In comparing opacity with the gasifier operating data, it appeared that particulate carry over to the boiler was most affected by flow of fresh limestone and gasifier load.

While the No. 3 burner nose plug was inserted all the way, manipulations of the No. 4 burner register were made in an effort to improve the No. 4 flame characteristics and to eliminate flame impingement on the waterwall adjacent to this burner. These efforts were futile, and flame impingement was eliminated only when the No. 3 burner nose plug was retracted and the burner itself put into service. Near the end of the run, the flame from burner No. 4 changed in shape and appearance, no longer being full and uniform but becoming more intense and more oriented toward the center of the boiler. It is believed that this was the result of a partial blockage of the burner by coke or tar.

Regenerator Operation

The regenerator was operated for approximately 4 hours during this run. Unfortunately, the duration of regenerator operation was so short that very few data were acquired. During operation, approximately 2 percent sulfur dioxide was produced in the regenerator. Operation of the regenerator was terminated when bed temperatures approached 1130°C (2000°F). When attempts were made to restart the regenerator, it was not possible to transfer material from the gasifier to the regenerator, restart was therefore impossible.

Sulfation

Gasification was discontinued at 22:18 on April 14, followed by sulfation which was initiated at 22:29 and appeared to finish at 23:12 which was surprising in view of the high sulfur loading on the gasifier bed. After sulfation ended the gasifier was put into combustion mode at about 500 kg/hr (11,200 lb/hr) oil rate to complete the burnout of duct and cyclones. It was shut down at 6:48 on April 15.

Inspection

After the gasification run, all equipment was shutdown and inspections of the gasifier, product gas cyclones and conduits, regenerator, and boiler were performed. The following observations were made:

- Gasifier--The main items observed during the gasifier inspection were coke deposits, coal needle and thermocouple sheath metal wastage and transfer slot blockage. There were several large hard masses, one almost 0.6 m (2 ft) in diameter. In addition, there were coke/bed material growths at several locations along the gasifier wall and in most of the pits. These agglomerations had been observed after previous runs; however, never as many or as large. The size and number of agglomerates was undoubtedly a result of below average stoichiometry levels and longer duration of the run.

The metal wastage noticed on the thermocouple sheaths was rather mild with slight flaking and discoloration of the surface. Attack on the coal needles, however, was more severe. There was insignificant flaking of the inner and outer surfaces on most of the coal needles, more severe on some than on others. In one case, needle No. 4, front, the needle was wasted through at the tip. This needle was removed for analysis and replaced with a spare. The metal attacks seemed more severe following this run, probably because of needle pluggage and subsequent overheating and also because of the longer duration of the run.

The transfer slot blockage was primarily caused by agglomeration of CaO , CaS , and coke which stayed in the transfer slots following regeneration shutdown. When this material was cleaned out following the run, it was found to be fairly hard and well packed, but quite easy to remove. All four slots were plugged.

Other observations made during the gasifier inspection were slag accumulations on the ceiling of the product gas conduits (resulting from lignite start-up), slight separation of the product gas conduits (resulting from lignite start-up), slight separation of thin layers of surface refractory on the gasifier walls and bed material accumulations on the floor of the product gas conduits (granular in nature with the surface material having an appearance similar to that of sulfated bed, while material below the surface had the appearance of unsulfated bed material mixed with coke). Thickness varied from 1.2 to 7.5 cm (0.5 to 3 in). Also, there was a crusty material on the cyclone walls (probably slag from lignite start-up), and a deposit of coke on the gasifier ceiling at the product gas outlet. Other than these deposits, the product gas conduits before and after the cyclones and the cyclones themselves were relatively clean and free from tar or coke. The "top hat" refractory plug was broken on cyclone A.

Oil leakage from the purge pipe surrounding the oil injector was discovered during the run. It was later determined that the tips of the oil guns were coking and forming a blockage. This malfunction forced oil back down the annulus, which contained purge gas, and ultimately produced the leakage through perforations in the outer pipe. Following the run, all the oil guns were removed (although some with great difficulty), cleaned and replaced. The failures in the outer pipe were also repaired.

- Regenerator--With the exception of the transfer slot blockage, inspection of the regenerator revealed no abnormalities. However, when the regenerator plenum was inspected, a puddle of oil was found. This indicated that the oil guns had leaked during the run causing oil to drip back through the fluidizing nozzles into the plenum.

Boiler

Inspection of the boiler revealed deposits on the boiler floor, waterwalls, and superheat tubes similar to those that had been seen following previous runs. The deposits on the floor were of a hard, crusty nature. The waterwalls and superheat tube deposits were about 0.4 to 0.8 mm (1/64 to 1/32 in) thick, crusty on the surface, with fine dust beneath. The tube deposits were loose and fell off when touched. It was noticed that not all the waterwall tubes had deposits on them; however, since there were piles of material at the base of some waterwall sections, it was assumed that deposits on these sections were dislodged during operation or cool down of the boiler. A section of refractory fell off from between burners Nos. 1 and 3. A block deposit, presumably carbon, was discovered on the boiler refractory surface separating waterwall tubes in the flame wash. This, and other observations made during the run, indicate significant tar and coke carry over from the gasifier to the boiler during the run. Apparently, all of the carbon carried over did not burn during product gas combustion, nor did carbon deposition burn during gasifier sulfation when the boiler was firing natural gas. There was no erosion of the boiler floor.

Emissions Data

Throughout the gasification run, effluent gases were continuously analyzed for oxygen, carbon dioxide, carbon monoxide, sulfur dioxide, and oxides of nitrogen from the boiler exit, the induced-draft fan exit, and the stack. Stack opacity was measured continuously and two EPA Method 5 particulate analyses were performed. The averaged data for each operating condition are contained in Table B-14. In addition, opacity readings for the run are shown in Figure B-7. It is apparent on perusal of Table B-14 that throughout the run oxides of nitrogen emissions were quite satisfactory, generally being lower when product gas was the only fuel used, with the exception of test periods, 7, 8, and 9. Sulfur dioxide emissions were fairly satisfactory through the fifth data set; it was then that the gasifier tripped. Before the trip, the SO₂ level had been increasing slightly, but after the trip the SO₂ level showed a significant increase. The as-read data are misleading because of dilution of SO₂ when natural gas is introduced as a fuel. Therefore, to give proper representation of the data, it was corrected for dilution. When the corrected data are observed, the increase in SO₂ following the gasifier trip appears more dramatic, but it does tie up with the increase in bed sulfur and the consequent reduction in the sulfur removal efficiency as can be seen later on in this chapter.

Particulate emissions were significantly lower during this run than in past runs, probably because of the decrease in fresh stone feed. Figure B-7

TABLE B-14. GASEOUS EMISSIONS DATA, RUN 3

Test period	Sample location	Oxides of Nitrogen			Sulfur Dioxide			Carbon monoxide			Corrected lb/10 ⁶ Btu	Opacity (%)
		Concentration (ppmv)	Standard deviation (ppmv)	ng/J	lb/10 ⁶ Btu	Concentration (ppmv)	Standard deviation (ppmv)	ng/J	lb/10 ⁶ Btu	Corrected ng/J		
1	BE ^a	96.6	6.4	64.5	0.150	115	39.3	107	122	0.285	21.8	6.6
	ST ^b	83.6	6.9	68.0	0.158	--	--	--	--	--	18.4	8.8
2	BE	106	31.3	70.1	0.163	247	46.7	228	335	0.780	9.1	6.2
	ST	82.0	--	64.1	0.149	215	--	234	345	0.802	20.0	7.0
3	BE	105	4.6	65.4	0.152	380	55.1	331	333	0.775	300.0	4.8
	ST	83.0	5.9	58.9	0.137	341	55.5	337	339	0.789	323.8	7.6
4	BE	96.0	5.7	63.7	0.148	683	159	628	--	--	10.0	5.8
	ST	54.0	19.8	44.3	0.103	533	117	606	--	--	520.0	8.7
5	ID ^c	96.3	2.5	70.1	0.163	423	31.2	434	--	--	136.0	7.2
	ST	75.0	--	64.5	0.150	390	--	469	--	--	40.0	10.2
6	BE	134	13.7	92.9	0.216	704	190	684	959	2.23	31.3	6.9
	ID	123	--	88.2	0.205	536	129	533	748	1.74	--	7.4
	ST	106	10.5	97.2	0.226	612	172	783	1,097	2.55	27.7	10.3
7	BE	160	14.9	108	0.250	587	96.9	533	856	1.99	22.5	6.5
	ID	149	2.1	102	0.237	484	133	462	748	1.74	32.5	6.9
	ST	110	--	93.8	0.218	465	--	550	886	2.06	25.0	9.6
8	BE	187	--	144	0.334	600	--	641	1,260	2.93	15.0	8.5
	ID	93	--	--	--	555	--	--	--	--	15.0	5.6
	ST	83.0	--	--	--	555	--	--	--	--	16.0	8.5
9	BE	160	15.2	104.7	0.242	670	59.3	606	929	2.16	21.6	6.0
	ID	156	7.8	106	0.247	668	113	636	972	2.26	19.6	6.8
10	BE	88.0	11.3	53.3	0.124	478	88.1	403	809	1.88	0.0	5.1
	ID	82.5	10.6	52.0	0.121	453	46.0	396	792	1.84	0.0	5.7

^aBoiler exit.^bStack.^cInduced draft fan exit.

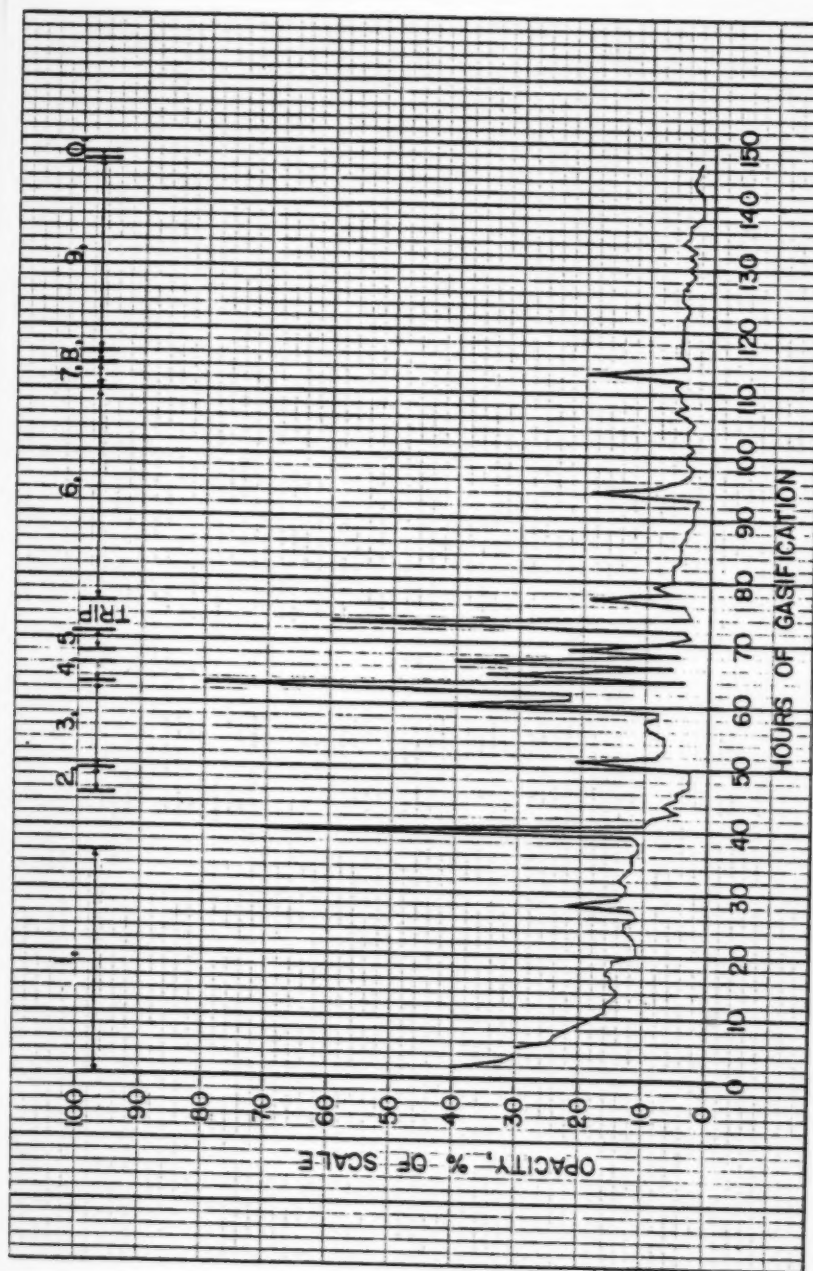


Figure B-7. Stack opacity, Run 3.

shows that opacity was reasonably low throughout the run and remained below 20 percent from the ninth hour on, with several exceptions. These exceptions coincided with the efforts to remove particulate matter deposition on tubes by suddenly increasing the natural gas (and hence the total combustion product flow) to the boiler.

The results of the two Method 5 analyses indicated exceptionally low particulate emissions of 17.6 and 22.8 ng/J (0.041 and 0.053 lb/10⁶ Btu input). These analyses were performed while product gas supplied 71 to 65 percent of total heat input to the boiler respectively. Correcting these data for dilution resulting from natural gas combustion gives 25.0 and 34.9 ng/J (0.058 and 0.081 lb/10⁶ Btu heat input), respectively.

The initial Method 5 test, performed during the 81st and 82nd hours of operation, resulted in an off-white (slightly yellowish) collection on the filter paper, very similar in nature to all the previous particulate test filters. However, the second Method 5 test, performed during the 135th and 136th hours of operation, yielded a black deposit on the filter paper, indicating a significant amount of unburned carbon going out the stack. The second test coincides with the change in burner No. 4 flame characteristics reported above.

Wet Chemical Analysis

On two occasions during the gasification run, gas samples were extracted from the induced-draft fan for analysis of oxides of nitrogen, and on three occasions for analysis of sulfur dioxide. These analyses were performed using EPA Method 7 for oxides of nitrogen and Method 6 for sulfur dioxide. The results of these tests and the corresponding readings from the continuous analyzers are shown in Table B-15. In all cases but one, the instrument reading was greater than the reference method, in one case by as much as 39 percent. However, in most cases, the continuous monitor was in fair agreement with the reference method.

Sulfur Removal

Shown in Figure B-8 is the sulfur removal efficiency throughout the run. Again, with the exception of a brief period, the regenerator was not in operation, and therefore the gasifier bed was replenished only by fresh limestone feed (which enters on the front side of the gasifier only). It is apparent that initially the sulfur removal was excellent, averaging approximately 88 percent during the initial 45 hours of gasification. During the next 25 hours of gasification, the SRE decreased from about 90 to 60 percent averaging 73 percent during this time period. The most radical reduction in SRE occurred after the 71st hour, when the gasifier tripped - at which time the gasifier bed temperature rose to almost 982°C (1800°F). Following the restart, bed material samples had a somewhat different appearance than before the trip. The post trip material appeared to be darker in color and more glazed (very shiny surface) than the pretrip material. Microscopic examination of the selected particles showed that the majority of the surface was smooth, exhibiting very few pores. This phenomenon undoubtedly contributed to the severe reduction in SRE.

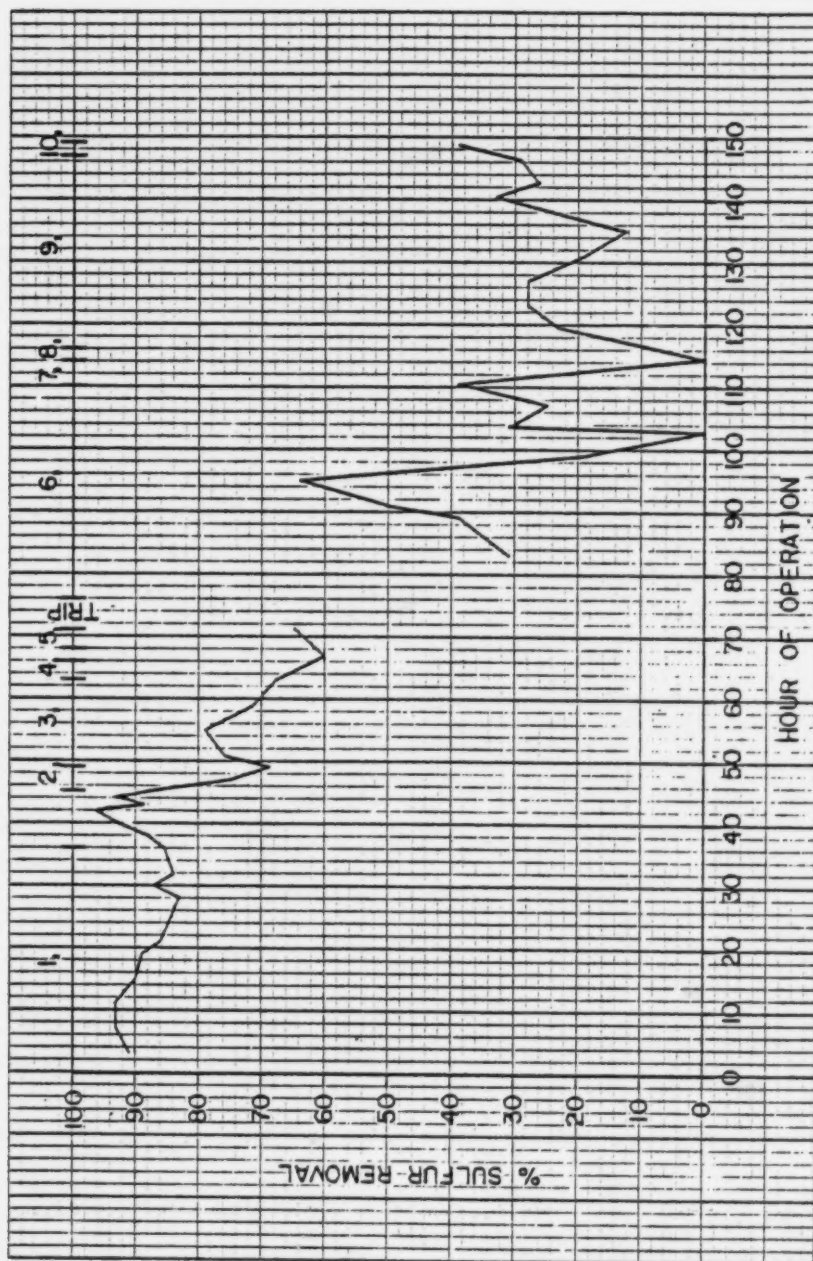


Figure B-8. CAFB sulfur removal efficiency, Run 3.

TABLE B-15. WET CHEMICAL GASEOUS ANALYSIS DURING RUN 3 (ppm)

Test Period	Gas Analyzed	Concentration Determined (ppm)	Average (ppm)	Standard Deviation	Continuous Analyzer Reading (ppm)
1	Sulfur Dioxide	86.6 85.5 86.6	86.2	0.6	119.5
2	Oxides of Nitrogen	96.21 98.66 95.23 96.09	96.55	1.47	115.6
2	Sulfur Dioxide	254 254 252.5	253.5	0.87	285
9	Oxides of Nitrogen	167.4 175.1 172.4 179.8	173.4	5.2	151.25
9	Sulfur Dioxide	723.6 720.0 792.4	745.3	40.8	575.5

Analysis of Gasifier Bed Material

Throughout the gasification run, samples of bed material were extracted every 4 hours from front and rear bed drains in the gasifier. All of these samples were analyzed for sulfur content, whereas every other set of samples was analyzed for carbon. Shown in Figure B-9 are the results of the sulfur analyses. It is apparent from observing these data that bed sulfur increased steadily from the outset of gasification until the 50th or 60th hour of gasification, after which the sulfur on the stone leveled out at approximately 22 percent. With the exception of a single 4-hour period, the regenerator was not in service. In addition, fresh limestone flow averaged 560 kg/hr (1240 lb/hr) (although at times it was as high as 1136 kg/hr (2500 lb/hr) or as low as zero). There does not seem to be any significant difference in sulfur content from gasifier front to rear, although the point where fresh stone enters the gasifier was approximately 1.22 m (4 ft) horizontally and 1.5 m (5 ft) above the front bed sample point.

Shown in Figure B-10 are the results of gasifier bed carbon analysis. The majority of the data fall in the 0.2 to 0.8 percent range, although

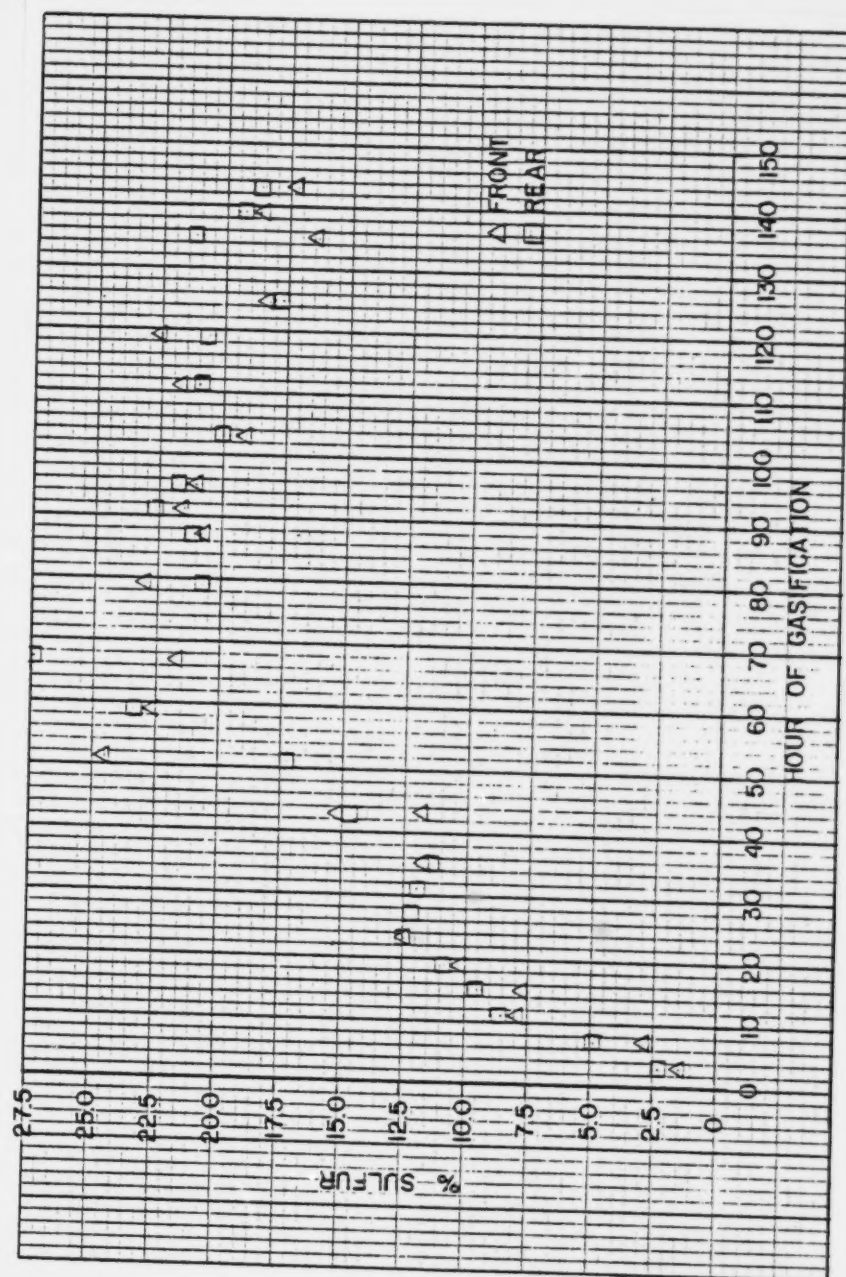


Figure B-9. Gasifier bed sulfur, Run 3.

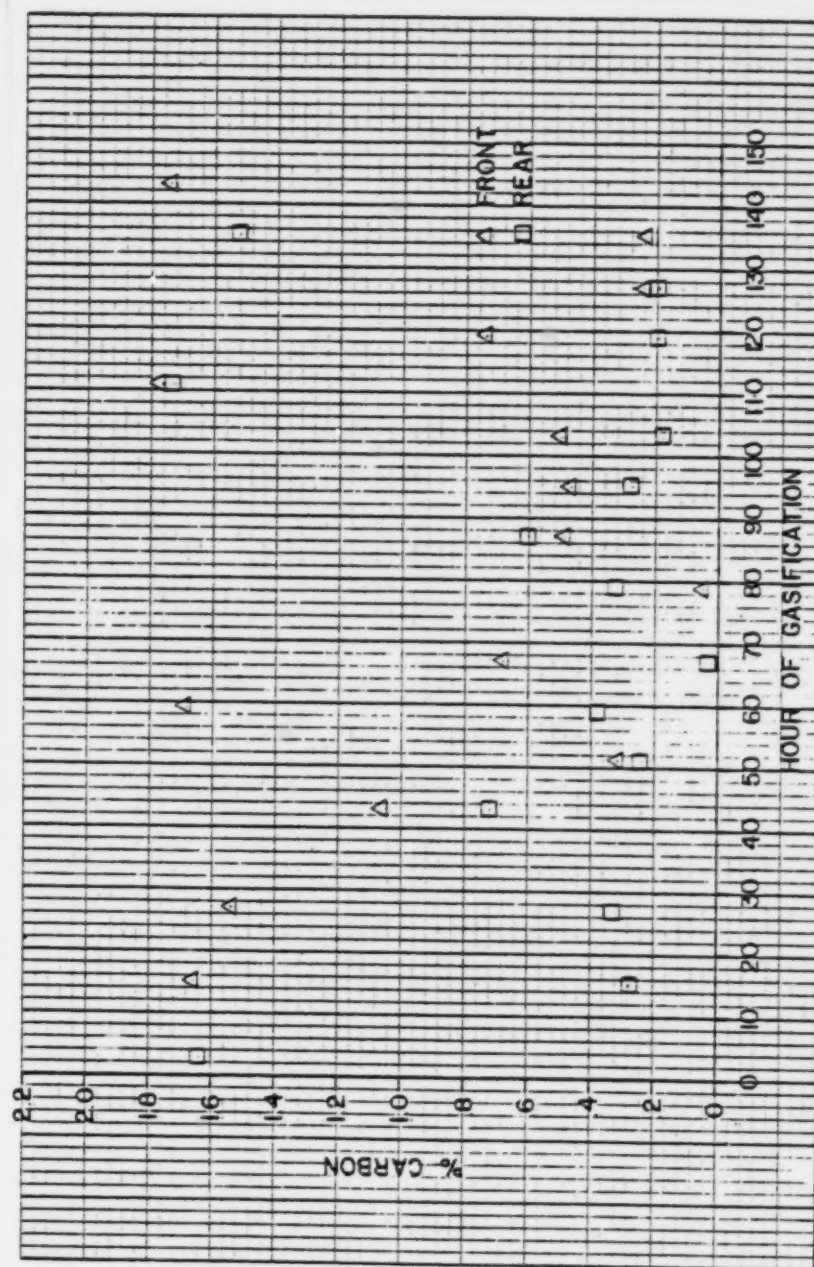


Figure B-10. Gasifier bed carbon, Run 3.

several of the samples (mostly from the front sample drain) are in the 1.5+ percent range. There appears to be no particular trend with these data.

Samples of gasifier bed material after sulfation were analyzed for remaining CaS and carbon because sulfation was unexpectedly short and "glazing" found on the bed material after the mid-run temperature excursion could have interfered with sulfation.

These analyses showed that:

S as CaSO_4 = 8%

S as CaS = 16%

C = 0.5%

The bed material was therefore drained and discarded in preparation for the next test run.

PART 4: FUEL OIL GASIFICATION (RUN 4)

Preparation for Run

Gasifier and regenerator solids were drained and dumped, boiler fines were removed and any refractories that needed repairing were repaired.

A number of work items were carried out either because they were necessary or deemed to improve the plant operation.

- Oil guns were pulled out, sleeves freed of oil and coke deposits and any necessary repairs were carried out.
- A nitrogen line was installed on the transfer sparger surge tank. This would allow the modulation of oxygen in the transfer gas or complete blanketing with an inert gas. A drain was fitted to the sparger surge tank, for draining off flue gas condensate, and also the transfer sparger solenoids were checked and repaired.
- An air line connection was installed to the coal distributor tables so that either flue gas or air would be available for pressurization of coal needles. Cortiflex bellows were installed on the actuator shafts to the coal distribution tables to prevent leakage.
- A new refractory plug was cast and installed on top of cyclone CY 101A.
- The Resox TM tail gas fan was eventually balanced after five balancing test runs were carried out, a specialist was called in and new floating and fixed bearings were installed.

- The main storage silo, containing wet lignite, was emptied, the lignite was air dried and replaced. A 30 cm (12 in) turbine ventilator was installed on top of the tank.
- The Lear Siegler SM-1000 and SM-800 sulfur dioxide analyzers and the Lear Siegler RM-7 opacity meter were repaired. These instruments were excellent while operational but their electronics were prone to failure and had caused some difficulty during previous test runs.

By the end of May most of the repairs and preparations were completed, 50.8 cm (20 in) of fresh limestone was placed in the gasifier and the pre-run check out was initiated. This was interrupted on June 5, 1980 when the motor of B101 main gasifier air blower failed: a piece of rotor, about 116 cm² (18 in²) was thrown off while the rotor was spinning at full revolutions, causing considerable damage. The out of balance forces caused a failure of the spindle bearing housing and a bent shaft as well as considerable external damage. The necessary spare parts were obtained and blower B101 was reassembled by July 11, and run preparations recommenced.

Start-up

Fresh limestone to a depth of 50.8 cm (20 in) was in the gasifier from the previous start-up attempt. This limestone was fairly coarse, similar to that which caused initial fluidization problems before Run 3.

Initial coal combustion was attempted on July 21, 1980 but failed due to malfunction of "A" weigh feeder and an unevenness of the gasifier bed. After leveling off the bed, the following day coal was fed again. This was done despite the low bed temperature. The attempt failed. On the third day, after soaking the bed with overhead burners the previous night, oil combustion was established.

The following 3 days were spent in building up bed level, trying to establish regenerator bed transfer using nitrogen as a transfer medium, and warming up the ResoxTM system.

On July 25, 1980 while draining the regenerator bed, the gasifier bed started to "free wheel" into the regenerator. Transfer slot temperature, (gasifier to regenerator), increased to 264°C (507°F) after about 1 minute. The indicated regenerator bed level increased slowly to about 66.0 cm (26 in) and the regenerator temperature (rear of bed) peaked out at 760°C (1400°F) and slowly dropped. The front of the regenerator was thermally inhomogeneous indicating poor mixing. After several attempts, continuous transfer was abandoned, and the regenerator bed slumped with the overhead burner in service to keep the refractory hot.

July 26, 1980: drained most of the regenerator bed out, put 2000 kg/hr (4400 lb/hr) of air to the regenerator plenum and pulsers on maximum. Within a few minutes there was a considerable bed transfer from the gasifier to the regenerator, the rear oil gun was placed in service in the regenerator and the

burner was switched off. Some more bed was drained out of the front of the regenerator and the front oil injector was put into service and regenerator temperature was maintained at about 870°C (1600°F) for a while with pulsers switched off.

Gasifier/Regenerator Transfer Tests

A number of tests were carried out by varying the pulsing rate with no overhead combustion or fuel oil injection into the regenerator; i.e., the regenerator temperature was maintained solely by transfer of hot lime from the gasifier.

Pulsing Interval Seconds	G/R Differential °C (°F)	Air to Regenerator kg/hr (lb/hr)	
3.7	170 (300)	2000	(4400)
2.5	140 (250)	2273	(5000)
2	128 (230)	-	-

Reducing the pulsing interval to 7 sec increased the differential rapidly and would have leveled out at about 260°C (500°F).*

At this stage all was ready to change to gasification, except for a short delay while the nitrogen supply was replenished.

Change to Gasification

July 29, 1980 by noon air had been backed off to give a near stoichiometric combustion. At 12:24 the conditions were:

Air and flue gas to plenum	11,950 kg/hr (26,310 lb/hr)
Plenum O ₂	11.5% vol.
Gasifier Temperature	850°C (1560°F)
Oil flow	545 kg/hr (1199 lb/hr)
Product Gas O ₂	1.7%

The oil flow was increased and by 12:27 the oil flow was 2918 kg/hr (6420 lb/hr) with product gas ignition at Nos. 3 and 4 burners with automatic compensation with natural gas. The actual transition from near stoichiometric combustion to gasification (with 8MW contribution from the gasifier) occupied about 15 seconds, after which there was only 2.5 MW contribution from natural gas in burner No. 2. There was no smoke observed at any point during the transition.

*The transfer rates implied by these differential temperatures were close to that measured in the cold model tests at FWEC Livingston, and the temperature differentials themselves were similar to values normally achieved by ERCA.

Gasifier Operation

During the more than 90 hours of gasifier operation in the fourth run, the gasifier provided from 62 to 100 percent of the heat supplied to the boiler. Throughout the run, the boiler was loaded at approximately 10.5 MW. Table B-16 shows average operating conditions of the gasifier during four periods of essentially stable operation. The first period of operation comprised the initial 34.5 hours of gasification. This period ended when, in an attempt to dislodge material from the boiler and the superheater tubes, the load was increased by an increase in the flow of natural gas. During the first period of operation, the oil flow to the gasifier varied from 2149 to 2442 kg/hr (4737 to 5383 lb/hr). This supplied approximately 63 percent of the heat input to the boiler (about 6.6 MW of the total 10.6 MW being produced). The gasifier stoichiometry during this initial period of gasification averaged 22.9 percent.

The second period of operation started about 4 hours after the completion of the first period and continued for 18 hours. During this period the boiler load was approximately 10.7 MW, with 6.5 MW (62 percent) supplied by the gasifier. Oil flow to the gasifier varied between 2386 to 2472 kg/hr (5260 and 5450 lb/hr) averaging 2440 kg/hr (5380 lb/hr). The gasifier stoichiometry averaged 21.3 percent during this period.

The third period of operation was the only period during which the boiler was fired with product gas alone. During this period, which lasted for 11 hours, oil flow to the gasifier varied between 3470 to 3602 kg/hr (7650 to 7940 lb/hr) (averaging 3524 kg/hr (7700 lb/hr)). The boiler load during this time was approximately 10.5 MW. The gasifier stoichiometry during this period averaged 19.4 percent. This operating period ended when the unit tripped (as a result of a problem in the ResoxTM plant). The unit trip will be discussed in greater detail in the section describing ResoxTM operation.

Following the unit trip, the gasifier was out of operation for approximately 7.25 hours. Gasifier and boiler conditions were allowed to stabilize for about 1 hour prior to the beginning of the fourth operating period. During this period, which lasted about 22 hours, the gasifier provided approximately 73 percent of the heat input to the boiler (7.7 of the 10.7 MW produced), with an oil flow which varied between 2533 to 2699 kg/hr (5585 to 5950 lb/hr) and averaged 2651 kg/hr (5845 lb/hr). Stoichiometry during this period averaged 22.8 percent. This operating period, (and, therefore, the test run itself) was terminated as the result of vibration problems with the ResoxTM tail gas fan and because of a decision to conserve No. 6 fuel oil. The gasification was therefore terminated by backing the gasifier temperature down to 846°C (1555°F) and switching over, on automatic control, from burning product gas to burning natural gas at 15:05 on August 2, 1980. The change over took approximately 1 minute and was followed immediately by backing off gasifier air to enter the sulfation mode. The gasifier temperature increase peaked at 920°C (1688°F) at 15:10, and was thereafter controlled at about 904°C (1600°F). Sulfation finished at 16:45 and was followed by a short period of combustion. Bed was slumped at 18:45 and isolated.

TABLE B-16. BOILER AND GASIFIER OPERATING DATA

	Operating Period			
	1	2	3	4
Boiler Load, MW				
Oil Flow to Gasifier, kg/h (lb/h)	10.5 2,278 (5,022)	10.6 2,428 (5,352)	10.5 3,524 (7,770)	10.7 2,568 (5,662)
Total Air and Flue Gas Flow To Gasifier, kg/h (lb/h)	12,400 (27,330)	11,900 (26,150)	13,020 (28,706)	11,700 (25,837)
Plenum Temperature, °C (°F)	164 (327)	158 (317)	142 (287)	158 (317)
Gasifier Bed Temperature, °C (°F)	870 (1,598)	861 (1,582)	863 (1,586)	863 (1,585)
Gasifier Freeboard Temperature, °C (°F)	828 (1,522)	829 (1,525)	841 (1,546)	831 (1,527)
Gasifier Stoichiometry, % % of Boiler Load Provided by Product Gas	22.9 63	21.1 62	19.4 100	22.75 73
Fresh Limestone Flow, kg/h (lb/h)	335 (738)	291 (641)	0 (0)	64 (140)

The following day (August 3, 1980) the gasifier bed was refluidized at 14:05 (bed temperature was 816°C (1500°F)), and placed in combustion mode after it had been slumped for 19 hours and 20 minutes. Regenerator was also refluidized and transfer initiated. After 1-1/2 hours both were slumped and allowed to cool.

Sulfur Capture

Sulfur removal efficiency (SRE) of the gasifier bed was determined throughout the run. Figure B-11 indicates that SRE was a fairly constant 80 percent for the first 22 hours of operation, but deteriorated over a 2-hour span to 70 percent. SRE then continued to deteriorate, reaching about 65 percent before suddenly jumping back to the 80 percent range. This sudden increase in SRE, which occurred in the 44th hour of operation, coincided with instrument calibration. After calibration the SO₂ instrument utilized to determine SRE was discovered to have been out of calibration by an amount approximately equivalent to the corresponding drop in SRE. Therefore, the SRE may actually not have changed at all (remaining in the 80 percent range). This calibration error was of the same order of magnitude as that previously experienced as a normal change in SO₂ concentration. Therefore, at the time of the SO₂ concentration change (resulting in a corresponding SRE change), there was no reason to suspect an instrument error.

During this period of operation, from approximately the 50th to 55th hour, the SRE again started to decline. This decline, which increased total sulfur loading on the gasifier, coincided with the start-up of ResoxTM. (The SO₂ analyzer utilized for SRE determination was checked and found to be in calibration.) Following the unit trip and subsequent restart, SRE increased once again to the 80 percent range and remained there to the end of the run.

Analysis of Gasifier Bed Material

Gasifier bed samples were taken from the front and rear of the gasifier at approximately 4 hour intervals. Various samples were then selected for analysis of sulfur (usually every other set of samples), carbon, and size distribution. Figure B-12 shows a plot of bed sulfur content versus time of gasification. During past runs increased bed sulfur was experienced with time; sulfur content exceeded 20 percent before peaking. An increase was not achieved in this run, however, the sulfur content of the gasifier exceeded 6 percent on only one sample. The primary explanation for the low percentage is that the regenerator was in operation for virtually the entire run, removing the sulfur from the system as SO₂.

Samples were periodically analyzed for total carbon and hydrogen, often with an additional analysis for carbonate carbon (carbon appearing in uncalcined material). The results of these analyses, shown in Figure B-13, indicate that there is no apparent trend to the carbon level. However, the total carbon can be significantly affected by carbonate carbon from uncalcined limestone. Unfortunately, an accurate determination of carbonate carbon is not easily made. Therefore, the total carbon indicated may not be a true indication of carbon.

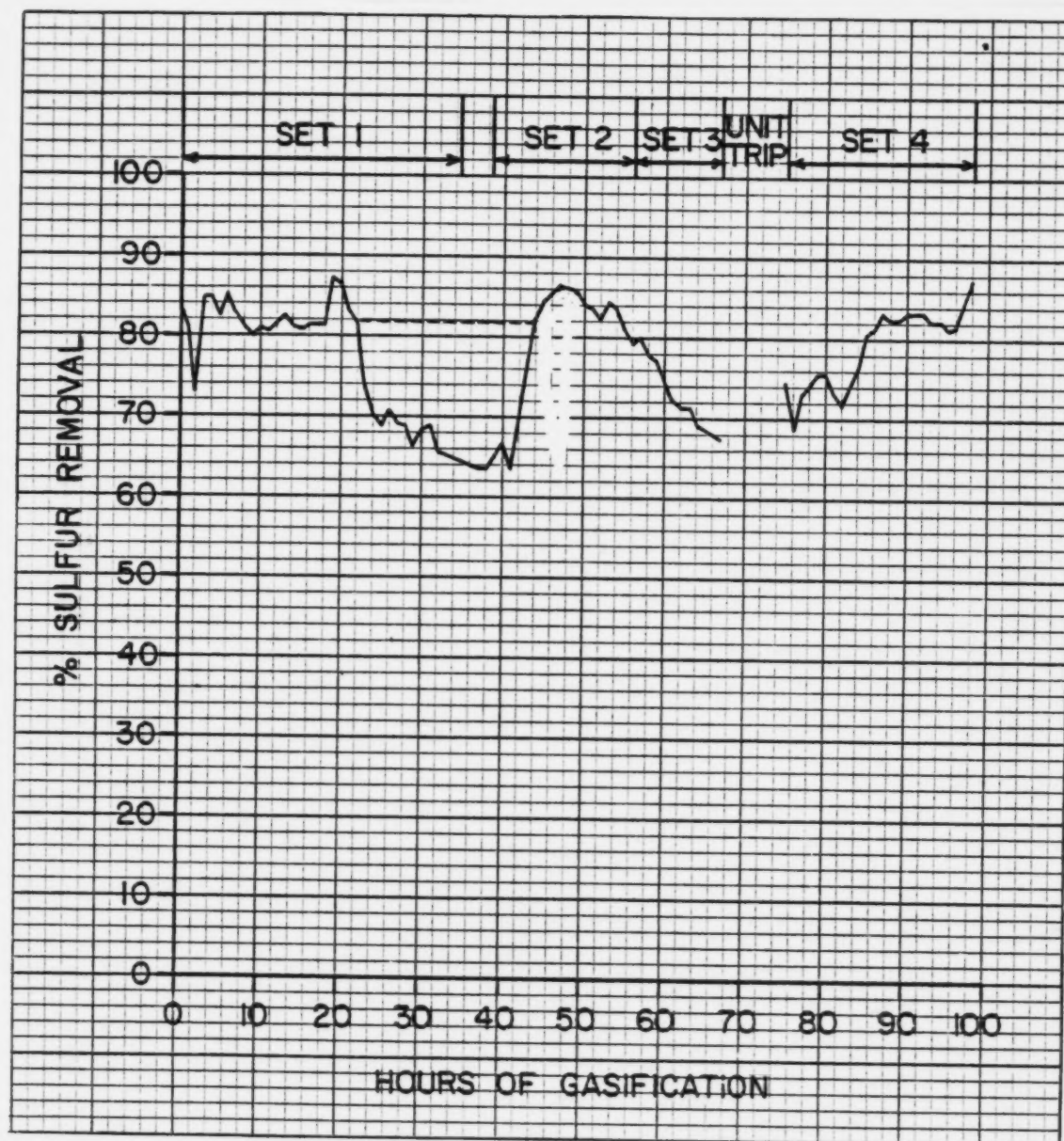


Figure B-11. Sulfur removal efficiency (SRE) of gasifier bed, Run 4.

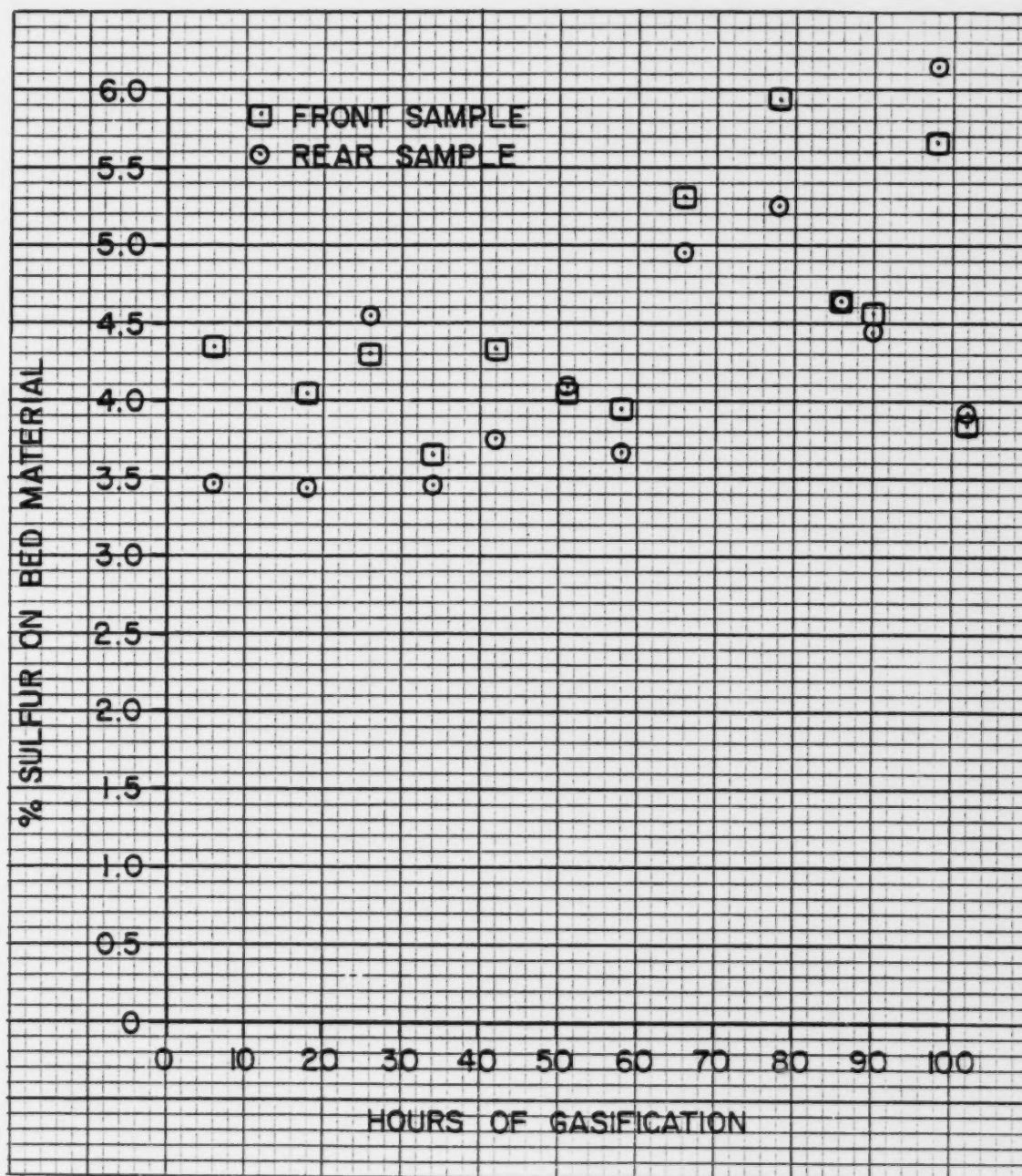


Figure B-12. Bed sulfur content vs. time of gasification, Run 4.

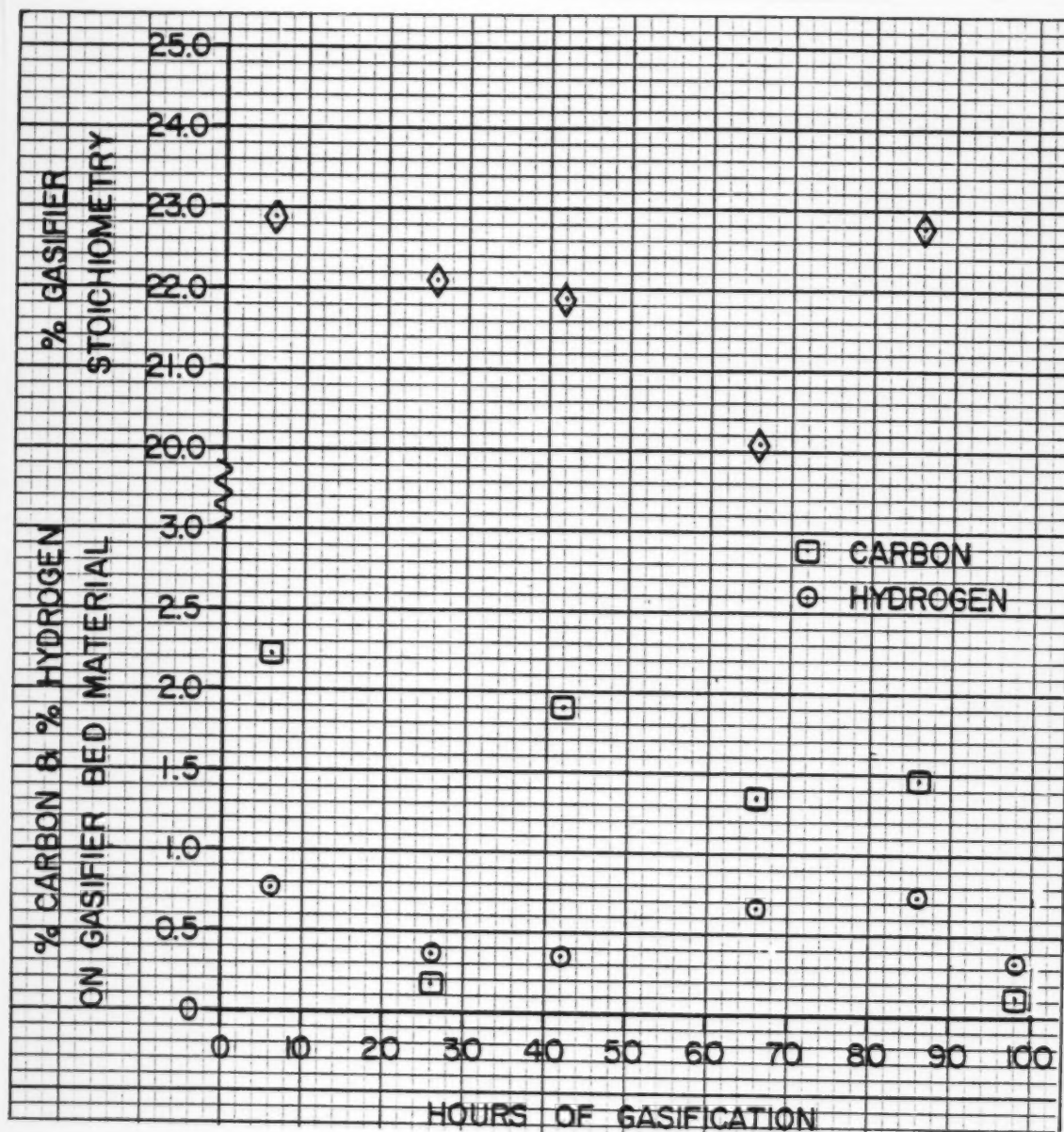


Figure B-13. Periodic analyses of gasifier bed samples, Run 4.

Sieve analyses of four samples, one at the beginning of the run, one at the end, and two in the middle, were performed to determine size distribution of the gasifier bed. These sieve analyses are shown in Table B-17. With the exception of the second sample, there was virtually no change in the particle size distribution. No explanation for the larger size distribution of the second sample can be offered at this stage of data analysis. All samples were obtained from the same location.

TABLE B-17. SIEVE ANALYSIS OF GASIFIER BED MATERIAL

Sieve size	% on screen for sample at indicated hour			
	Hour 6	Hour 26	Hour 66	After sulfation
8	11.5	29.2	10.2	10.1
10	10.6	16.6	9.4	10.5
12	11.9	16.3	10.9	11.7
14	13.2	13.5	12.5	13.1
16	10.2	7.3	9.9	10.0
20	23.3	11.0	23.2	22.1
30	14.6	4.6	17.5	16.4
Pan	4.7	1.5	6.4	6.1

Regenerator Operation

During this run the regenerator was operated as a regenerator or sulfator for essentially the entire run (97 hours 40 min). Operation as a sulfator occurred, primarily, in the early stages of gasification and for a short period after the system trip. For the remainder of the operating time the unit was operated in the regeneration mode. Unfortunately, the instrumentation used for determining SO_2 in the regenerator off-gas (a Lear Siegler SM-800 SO_2 analyzer designed for high SO_2 concentrations and high temperatures) was out of service (a typical condition for this particular instrument). Therefore, continuous determination of SO_2 in the regenerator off-gas was not possible. However, occasional samples of regenerator off-gas were analyzed by a wet chemical procedure. Results of these wet chemical analyses and the results of continuous analyses of O_2 , CO , and CO_2 are shown in Table B-18.

Samples of regenerator bed material were extracted at approximately 4 hour intervals. One sample from each 24-hour period was analyzed for carbon and hydrogen, while one sample at 12-hour intervals (every third sample) was analyzed for sulfur. The results of these analyses, shown in Table B-19, indicate that carbon levels remained fairly low, much lower than bed carbon levels in the gasifier at corresponding times of operation. The regenerator bed sample drain is located near the transfer slot, which transfers material from the regenerator to the gasifier. Apparently not all of the incoming carbon undergoes reaction in the regenerator before the return of the material

TABLE B-18. REGENERATOR OFF-GAS ANALYSIS

Hours of Gasification	Sulfur Dioxide (vol %)	Oxygen (vol %)	Carbon Monoxide (vol %)	Carbon Dioxide (vol %)
5	2.13	0.45	0.085	3.8
6	1.78	-	0.060	3.8
7	2.72	-	-	-
8	2.25	-	-	-
27	0.95	-	-	-
30	1.24	-	-	-
31	1.48	-	0.040	2.9
33	1.24	-	-	-
54	3.19	1.2	0.035	2.9
56	2.96	1.0	0.040	3.1
60	4.91	1.0	0.040	4.2
64	4.44	1.1	0.070	2.1
80	1.77	-	0.045	4.6
92	2.0	-	0.105	4.6

TABLE B-19. ANALYSES OF REGENERATOR BED MATERIAL

Time of Gasification (hour)	Total Sulfur (wt %)	Total Carbon (wt %)	Hydrogen (wt %)	Gasifier Sulfur (wt %)
6	3.19	0.16	0.53	3.48
18	2.64	-	-	3.44
30	2.46	0.11	0.28	4.01
42	2.51	-	-	3.76
54	1.92	0.42	0.32	3.88
66	2.28	-	-	4.95
78	4.88	0.29	0.30	5.26
90	3.86	-	-	4.45
Post Sulfation (+ 4 hours)	3.96	0.44	0.34	3.85

to the gasifier. When compared with gasifier bed sulfur for the same period indicated in the last column, the regenerator bed sulfur levels indicate an obvious reduction in bed sulfur.

Three samples of regenerator bed material, one at the beginning, one at the middle, and one at the end of the run, were analyzed for size distribution. The results of these analyses are shown in Table B-20. These analyses show that the regenerator bed material had essentially the same size distribution as the gasifier bed material.

This was the first occasion on which the regenerator was operated for an extended period of time. Prior to this run, transfer of material between gasifier and regenerator was accomplished by utilizing flue gas (containing approximately 7.5 percent oxygen and 8 percent water) as a transport medium; thus transfer slot plugging was a chronic problem. During this run, however, the transport medium utilized was nitrogen, which did not cause plugging of transfer slots. The only drawback was the excessive consumption of nitrogen.

ResoxTM Operation

ResoxTM was operated for the first time during this run. ResoxTM operation was initiated in the 51st hour of gasification and continued until a unit trip terminated the operation approximately 17 hours later. After all systems were put back on line an attempt was made to restart ResoxTM. This effort was futile as the ResoxTM tail gas fan had severe vibrational difficulties (as much as 26 mil). The test run was terminated before additional efforts to restart ResoxTM could be made.

The quantity of sulfur recovered during ResoxTM operation is not known, but a conservative estimate is approximately 908 kg (2000 lb). This estimate was based on the approximate sulfur (as SO₂) input to ResoxTM and a conversion of 70 percent of the SO₂ to sulfur.

ResoxTM Unit Trip

After approximately 68 hours of operation, a trip of the ResoxTM unit occurred. As a result, the regenerator, gasifier, and boiler tripped. An analysis of the events leading up to and immediately following the ResoxTM trip lead to the following conclusions concerning causes and effects of the trip.

Prior to the ResoxTM trip, difficulties were encountered in controlling ResoxTM temperatures. A slight temperature excursion resulted. This increase in temperature caused a corresponding increase in sulfur temperatures and, consequently, sulfur viscosity. The increase in sulfur viscosity caused the sulfur pump motor to overload and trip on high current draw. The pump motor trip, in turn caused a ResoxTM trip. In the event of a ResoxTM trip, a valve isolating ResoxTM from the regenerator closes and a vent valve opens, venting regenerator off-gas to the stack. Because the ResoxTM isolation valve closed quickly and the regenerator vent to the stack opened slowly, the overhead pressure in the regenerator increased to the trip set point, whereupon both the regenerator and gasifier tripped. Since at the time, the boiler was being

TABLE B-20. SIZE DISTRIBUTION OF REGENERATOR BED MATERIAL

Sieve No.	% Retained on screen for a sample at the indicated hour		
	Hour 6	Hour 54	After Sulfation
8	9.6	10.2	9.7
10	8.5	9.0	10.3
12	9.5	10.0	12.2
14	12.6	11.6	13.3
16	9.5	9.5	10.7
20	25.4	23.7	22.7
30	18.2	18.8	15.9
Pan	6.7	7.2	5.2

fired solely with product gas, the boiler tripped also. Vibration problems with the tail gas fan prevented ResoxTM from being put back into operation.

Boiler Operation

Throughout the gasification run, boiler operating parameters were monitored and recorded. The pertinent operating data for each run period are shown in Table B-21, along with boiler efficiencies calculated by loss method and by input-output method. Both methods are included even though efficiency calculation by loss method is deemed more accurate. Past input-output determinations have given erratic results and there were some questionable flow measurements. In general, boiler efficiency appeared to improve with increased quantities of product gas. This improvement was the result of a decrease in moisture from combustion. The exception to this finding occurred during the final operating period when stack gas temperatures were lower than had yet occurred in the run, resulting in a lower dry flue gas loss.

Emissions Data

Throughout this run, samples of effluent gas from the boiler exit, induced draft fan exit, and the stack were continuously drawn and analyzed for O₂, SO₂, NO_x, and CO₂. Averages of the data recorded during each of the operating periods are shown in Table B-22. A review of the data in this table reveals that a small amount of air leaked into the boiler exit samples. Oxygen values in the boiler exit samples were therefore higher than they should have been (higher than some of the O₂ values in the ID fan exit samples) and other gaseous concentrations were lower than they should have been (as a result of dilution). However, when mass flows of constituents are determined, the effects of dilution are eliminated.

As Table B-22 demonstrates, NO_x emissions were fairly low (as they had been in previous runs), not exceeding 90.4 ng/J (0.21 lb/10⁶ Btu). The lowest NO_x emissions, averaging 64.5 ng/J (0.15 lb/10⁶ Btu), were experienced during the third operating period in which the boiler was fired with product gas along (no natural gas).

SO_x emissions throughout the duration of the run were quite low, since approximately 80 percent of the fuel sulfur was removed during the operating periods (Figure B-11). The third operating period, when ResoxTM was in service, was exceptional in that sulfur removal continuously declined to a low of about 65 percent (80 percent SRE corresponds to an emission of approximately 258 ng/J (0.6 lb/10⁶ Btu), while 70 percent SRE corresponds to approximately 387 ng/J (0.9 lb/10⁶ Btu)). The failure to remove any fuel sulfur would result in an emission of approximately 1290 ng/J (3.0 lb/10⁶ Btu).

Carbon monoxide emissions were low throughout the run. The highest concentrations of emissions were experienced during the third operating period, when the boiler was fired with product gas only. This was not surprising since the product gas flame was long and lazy, especially when compared with the natural gas flame. The product gas flame filled virtually the entire

TABLE B-21. BOILER OPERATING DATA

	Operating Period			
	1	2	3	4
Boiler Load, MW	10.5	10.6	10.5	10.7
Natural Gas Flow, m ³ /h(ft ³ /hr)	1,455 (50,730)	1,147 (40,000)	0	1,046 (36,500)
Oil flow to Gasifier, kg/h(lb/hr)	2,283 (5,022)	2,433 (5,352)	3,532 (7,770)	2,574 (5,662)
% of Boiler Load Supplied by Product Gas	63	62	100	73
Steam Flow, kg/h(lb/h)	49,880 (109,730)	54,430 (119,750)	48,940 (107,670)	49,770 (109,500)
Boiler Efficiency, Output/Input, %	83.49	74.47	84.11	86.59
Boiler Efficiency, Loss Method, %	83.93	83.72	85.15	85.94

TABLE B-22. GASEOUS EMISSIONS DATA, RUN 4

Test period	Sample location	Oxides of Nitrogen			Sulfur Dioxide			Carbon monoxide concentration (ppmv)	Carbon dioxide concentration (Vol %)	Oxygen concentration (Vol %)	Opacity (%)
		Concentration (ppmv)	Standard deviation (ppmv)	ng/J	1b/10 ⁶ Btu	Concentration (ppmv)	Standard deviation (ppmv)	ng/J	1b/10 ⁶ Btu		
1	BE ^a	129	7.6	90.4	0.21	149	24.4	146	0.34	6.82	
	ID ^b	134	7.4	90.4	0.21	129	41.9	120	0.28	6.43	6.9
	ST ^c	105	10.1	86.1	0.20	436	109.6	490	1.14	8.77	
2	BE	117	11.0	81.8	0.19	124	10.3	120	0.28	6.72	
	ID	119	10.5	81.8	0.19	125	10.3	120	0.28	6.68	4.4
	ST	100	3.9	81.8	0.19	506	160	580	1.35	8.95	
3	BE	100	11.0	73.2	0.17	320	57.5	327	0.76	7.01	
	ID	92.2	29.8	64.5	0.15	316	54.5	314	0.73	6.66	4.5
	ST	82.1	5.1	68.8	0.16	338	189	404	0.94	9.13	
4	BE	111	6.1	86.1	0.20	179	39.9	197	0.46	8.43	
	ID	119	4.5	90.4	0.21	217	42.8	223	0.52	7.58	2.6
	ST	94.2	3.3	86.1	0.20	599	58.2	744	1.73	9.77	

^aBoiler exit.^bInduced draft fan exit.^cStack.

furnace area, sometimes even extending into the convection passes. The CO levels during this period, however, were relatively stable in the 170 to 190 ppm range, with an average of 176.5 ppm.

Stack opacity was continuously measured and periodically recorded. The opacity during the run versus time of gasification is plotted in Figure B-14. The opacity was consistently below 20 percent and was less than 10 percent for most of the run. Exceptions to these figures include the 34th hour of gasification, when opacity went to 100 percent, and the 37th hour, when it went to 80 percent. During these periods of time the boiler load was increased by the introduction of more natural gas in an effort to dislodge dust deposits from boiler tubes, superheater tubes, and air heater tubes. Whenever this procedure was enacted, large quantities of dust were dislodged and, in general, stack opacity was increased.

Inspection

Gasifier--

A post-operation inspection of the gasifier uncovered no unusual effects. Additional corrosion of the coal needles was evident and small pieces of refractory material had been dislodged from the walls and floor of the vessel. The only problem requiring repair work was damage to the end of the gasifier division wall. Examinations made at the conclusion of prior runs revealed that some of the tongue and groove brickwork at the end of the wall in the first 0.91 m (3 ft) of height had begun to separate. Although attempts were made to repair these separations, a section of wall approximately 15 x 60 cm (6 x 24 in) fell away. Repairs were made with coral plastic refractory.

Although the transfer slots between the gasifier and regenerator were full of bed material, post-operation inspection revealed that this material was granular and easily removed. Flow through the transfer slots was therefore not obstructed.

Regenerator--

A post operation inspection of the regenerator uncovered no ill effects. The normal chipping and erosion of refractory on the regenerator was observed.

Gasifier Cyclones and Conduits--

Both gasifier cyclones and both conduits from the cyclones to the burners appeared to have suffered no adverse effects. A small amount of material, presumably fines from the bed or coal ash from start-up was found on the walls of the cyclones and on the product gas conduits.

The conduits from the gasifier to the cyclones had an accumulation of bed material of nearly normal size. The amount of material on the conduit floor was shallow at the entrance of the conduit, increased in depth to about 10 to 15 cm (4 to 6 in) over the first 1 m (3 ft) of conduit, and then leveled off at a depth of approximately 10 cm (4 in) throughout the rest of the conduit. Both conduits showed similar bed material accumulations. This material was cleaned out in preparation for the next run.

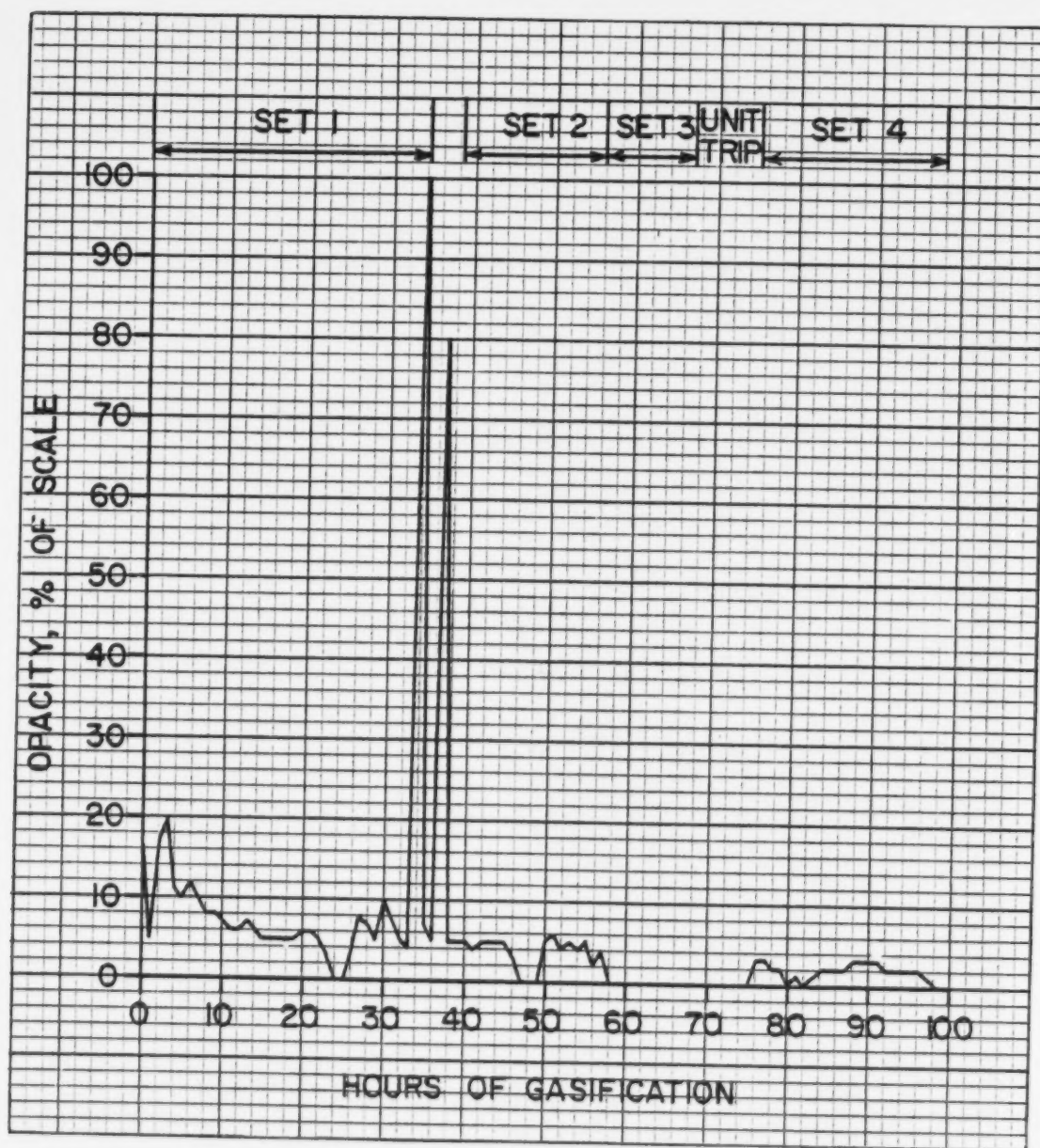


Figure B-14. Opacity during run vs. time of gasification, Run 4.

Boiler Tube Deposits

Tube deposits were similar to those encountered in previous test-run inspections: a very thin crust (approximately 0.04 cm (0.016 in) thick) developed on some tubes, while others had a very thin layer of light dust. The quantity of dust on the waterwalls was exceptionally light, lighter than had been experienced in past runs. This circumstance may have been misleading, however, as there was approximately 2.5 cm (1 in) of material on the floor, suggesting that a large amount of flakes had fallen from the tubes during operation or cool down. This material was easily removed from the tubes by scraping or light brushing.

Burner Deposits

Deposits of lime containing small amounts of slag were observed on both low-Btu burner throats. This material was peeling off the surface of the refractory and was easily dislodged.

The tips of the natural gas guns had carbon deposits on the protruding surfaces. The deposit on the No. 3 gun was heavier than that on the No. 4 gun.

Furnace Floor

Accumulations of debris on the furnace floor were as much as 2.5 to 5.0 cm deep (1 to 2 in) in some areas. The greatest accumulations were located on the sides and rear of the furnace (a result of tube deposits falling off and of flames pushing material from the area in front of the burners). The floor in front of the burners, which had suffered erosion damage and been repaired following the second run, was eroded in a somewhat similar fashion. But the erosion was not as severe and no repairs were necessary.

Superheater Tubes

The particulate deposition on the superheater tubes was significantly lighter than previously experienced. During previous runs material had built up on the leading edge of the superheater tubes in a ridge-like fashion. Accumulations at the base of the superheater tubes were also extensive, up to 18 to 20 cm (7 to 8 in) deep. As with dust accumulations in the furnace area, large amounts of material had probably deposited on the tubes and then fallen to the floor at the base of the tubes during operation or boiler cool down.

Operational Problems

During the run several problems were experienced:

- Repeated blockages of cyclone drain legs, particularly on CY 101A.
- Overload trips on the cyclone drain rotary valves.
- Solids cooler tripped out again, this time the trip was caused by the rotary valve at the bottom of the spent solids baghouse.

- Regenerator cyclone drain valves caused some problems, mostly gland leakage and failure to sense the end of stroke.
- Vibration in ResoxTM tail gas fan.
- Blockage of coal needles during the start up period.
- During the run a fire started in the main lignite silo and required the use of the water deluge system. Subsequently, the lignite had to be removed, dried and replaced.

References

1. Bazan, John A. Chemically Active Fluid Bed (CAFB) Process Solids-Transport Studies. Prepared by Foster Wheeler Energy Corporation, Livingston, New Jersey, for the U.S. Environmental Protection Agency. EPA-600/7-77-144. October 1977.

APPENDIX C

PART 1: OIL AND LIGNITE GASIFICATION (RUN 5)

Preparation for Run

The gasifier, regenerator and boiler were cleared and the partition wall between the front and rear parts of the gasifier was repaired. Considerable effort was expended in further attempts to balance the ResoxTM tail gas fan B-201 and to improve and brace its foundations. Coal injectors were shortened to alleviate the plugging problems.

First Start-up Attempt

Sixty-nine cm (27 in) of fresh limestone was preheated with overhead burners and flue gas on September 9, 1980 in preparation for start-up of Run 5. At 19:05 the gasifier bed was fluidized with flue gas and 1580 kg (3490 lb) of lignite fed in but the start-up had to be aborted at this stage because both spent fines lines and the regenerator spent fines line became blocked.

On September 10 the fines lines were cleared. The gasifier bed was refluidized using 22,700 kg/hr (50,000 lb/hr) of mostly flue gas and a further 1810 kg/hr (4,000 lb/hr) of lignite were injected over a period of 1 hour, at the end of which time the gasifier temperature was rising steadily and preparations were being made to inject oil. An interruption, in the form of smoke from the spent fines bag house, caused a shutdown. A nitrogen blanket was used to extinguish a fire in the spent fines bag house. Start-up was delayed until some necessary repairs were carried out.

Second Start-up Attempt

On September 22, 1980, a second attempt was made to start Run 5. Initial fluidization of 50.8 cm (20 in) of limestone containing some lignite was successfully carried out at 16:22 and a conventional start-up similar to that described previously, followed using lignite as the initial fuel. Oil was first injected at 17:42 and by midnight the gasifier had settled on steady combustion and the bed level had been built up to a fully calcined 53 cm (21 in).

On September 23, 1980 there was steady combustion all day. Gasifier bed had been built up to 74 cm (29 in). There were two minor problems: the store feed motor required a new bearing and the flue gas recycle baghouse auger needed a cleanout. The regenerator was being preheated.

On September 24, 1980 there was steady combustion in the morning. Regenerator was started at about 9:00 by fluidizing and starting pulsers. By 10:00 the temperature in the regenerator was high enough to prepare for oil injection which was initiated at 10:04 at a temperature of 718°C (1325°F).

Change to Gasification

By mid-morning on September 24, 1980, the gasifier had settled at 885°C (1625°F) with 86 cm (34 in) of bed and burning about 450 kg/hr (1000 lb/hr) of oil using 12,160 kg/hr (26,800 lb/hr) of mostly air in gasifier plenum. Total air/flue gas flow to the regenerator was 2050 kg/hr (4520 lb/hr) with sufficient oil to the regenerator to maintain a temperature of about 870°C (1600°F). At 11:05 most of the air to the plenum was replaced by flue gas, air was put onto the coal tables (empty at this stage) and a nitrogen purge was put onto the start-up burners. Pilots were lit in both burners No. 3 and No. 4 and there was natural gas in burners No. 1 and No. 4 firing at 11 MW. With 11 oil guns in the gasifier at 11:30, all oil was taken off the regenerator injectors and a temperature of 746°C (1375°F) was maintained by bed transfer only.

At 11:40 the oil rate was increased and by 11:41 gasification at an oil rate of 2840 kg/hr (6270 lb/hr) had been initiated. At 11:48 regenerator temperature had reached 843°C (1550°F) and was rising rapidly, indicating that regeneration was about to start: in fact it started by 12:10.

Gasification

From noon on September 24, 1980, oil gasification settled to a steady rate of a little under 2,300 kg/hr (5,000 lb/hr) through 14 oil guns at a gasifier temperature of 840°C (1550°F) and regenerator temperature of 980°C (1800°F). The boiler operated at 10.5 MW with one-third of the heat supplied as natural gas via No. 1 burner and the remaining heat being supplied as product gas via No. 3 and 4 burners.

A change in operating conditions was made at 16:33 on September 25 when the regenerator tail gas (which hitherto was being vented to the stack) was diverted to the ResoxTM plant and the ResoxTM tail gas was vented via fan B-201 to the gasifier flue gas recycle intake. At the changeover time, the regenerator off gas contained 2.3 percent SO₂ by volume and the regenerator was running at 982°C (1800°F) with pulsers cycling 1 second on and 4 seconds off using nitrogen.

On September 26, 1980, the initial conditions were essentially the same as during the previous day but from about 9:30 onward a few problems recurred. At 9:30 the ResoxTM tail gas fan had to be taken out of service due to severe vibrations so the regenerator tail gas was backed out of ResoxTM and once again vented to the stack. The planned operation at a higher oil rate had to be abandoned because of delays in delivery of No. 6 fuel oil. At 11:30 the oil rate was backed down a little instead to conserve oil. At about 18:00 a blockage was noticed in the CY 101A dip leg but was cleared quite easily by rodding and was in service by 18:40.

An attempt was made at 20:32 to place ResoxTM tail gas fan back in service and pass the regenerator off gas through the ResoxTM plant. This attempt was abandoned after 8 minutes with 0.66 mm (.25 in) vibration recorded at the tail gas fan.

There were no problems with the gasifier, the regenerator, or the boiler.

On September 27, 1980, with gasifier and regenerator both operational (regenerator venting to stack) and the ResoxTM plant on hot stand-by, the only problem appeared to be the ResoxTM tail gas fan. After another attempt at balancing, the tail gas fan was run up to speed at 9:15 and then run down again because vibrations of 0.61 mm (.24 in) were being registered.

At this stage, a decision was made to run ResoxTM on a pressure balance system with the flue gas recycle fan B102 providing the extraction suction for the ResoxTM plant and the ResoxTM tail gas fan B201 free-wheeling and inactive. At 9:30, therefore, the regenerator off gas was once again lined through to the ResoxTM plant and by 10:52 the sulfur loop was in service and the plant appeared to be operational. The next operational change was made at 14:15 when the flue gas to the regenerator was reduced and air to the regenerator was increased resulting in a rapid increase in the regenerator temperature from 982°C to 1066°C (1800°F to 1950°F) with a consequent increase in the regenerator off gas SO₂ level. The condition soon settled at 2,360 kg/hr (5,200 lb/hr) oil and 11,400 kg/hr (25,200 lb/hr) air and flue gas to the gasifier, which was running at 870°C (1600°F). The regenerator was running at 1066°C (1950°F) producing about 4 percent SO₂ by volume which was being fed to the ResoxTM process. These conditions were maintained through the night with only one interruption at 19:55 caused by a ResoxTM drum low level alarm causing a ResoxTM plant trip. The regenerator vent to stack was opened and ResoxTM vent to B102 was closed, the ResoxTM plant fault was rectified and venting the regenerator to ResoxTM was resumed within minutes.

On September 28, 1980, gasifier and regenerator were operating steadily but at 3:40 a high pressure alarm at the ResoxTM inlet necessitated venting the regenerator off gas to the stack to prevent a general plant trip caused by high regenerator-to-gasifier differential pressure.

Since ResoxTM operation was on a pressure balance, a solution to this problem had to be found without the use of the ResoxTM tail gas fan. Increasing water flow to the regenerator off gas cooler reduced the ResoxTM inlet temperature to below 538°C (1000°F) and caused a reduction in the ResoxTM pressure drop so that by 8:00 the regenerator vent to the stack was closed again and the ResoxTM plant was once again operational. At 9:10 a further improvement in the situation was achieved by backing off some flue gas from the regenerator (fluidizing velocity was reduced from 1.25 to 1.1 m/sec or from 4.1 to 3.6 ft/sec) in order to reduce the total gas flow through the ResoxTM plant. This reduction in flow through the regenerator also had the effect of increasing the regenerator off gas SO₂ level from 4 percent to near 7 percent.

GCA/Technology personnel were about to start the sampling test. Therefore, at 10:50, transfer to 100 percent oil operation was initiated. By 11:10 all natural gas was out of the boiler which was running at 9.5 MW on product gas from 2,495 kg/hr (5,500 lb/hr) of oil. Sampling tests commenced soon after noon and were completed at 22:15 at which time natural gas was introduced, at a low rate, into burner No. 1, igniters were put into service on burners No. 3 and 4 and oil rate was reduced by 113 kg/hr (250 lb/hr) to maintain boiler load at 9.5 MW.

On September 29, 1980, at 9:00 all natural gas was isolated from the boiler to allow more sampling by GCA/Technology. Oil rate was not increased to compensate so boiler output dropped to 8.4 MW. Because of a slight stack plume, pilots were put in service again on No. 4 at 12:00 and on No. 3 at 12:50. They were taken out of service again at 13:04. Sampling was completed by 20:25 but the plant remained on oil only operation until the next day.

At 21:10 a high pressure drop across the flue gas baghouse was noticed, 5.2 kPa (20 in H₂O). This problem had been building up slowly but at this stage had reached a level at which the flow of flue gas recycle was being affected.

On September 30, 1980, gasification with regeneration and ResoxTM conversion of sulfur continued through the night, with no natural gas in the boiler, until 10:05 when a small rate of natural gas was put through the No. 1 burner and pilots were lit on product gas burners No. 3 and 4. The main reason for caution was the increase in pressure drop across the flue gas baghouse which was forcing a departure from the previous mode of operation and could possibly have forced a shutdown.

At 14:55, the recycle gas damper to the gasifier reached a wide open position to maintain the necessary flow of air and flue gas mixture. Conditions appeared to have stabilized at this point and were maintained until the next day with two noncritical aspects:

- At 17:20 a faulty pump seal on one of the ResoxTM sulfur pumps caused some pressure surges as far back as the regenerator; however, the problem was dealt with before a plant trip occurred and steady operation was restored by 17:30. Boiler operation was not affected.
- At 23:50 a low coal level in the ResoxTM plant caused some rapid coal handling and upset the temperature profiles in the ResoxTM plant for a while. Boiler operation was not affected.

October 1, 1980: since about noon on the previous day the air rate to the gasifier had been reduced as the availability of flue gas decreased, to maintain plenum oxygen at about 12.5 percent and the gasifier temperature at 900°C (1650°F).

Early in the morning, a decision was made to attempt partial replacement of oil with lignite before a forced shutdown.

Lignite and Oil Combined Gasification

At 9:30 on October 1, 1980, the lignite needles were rodded clear and the lignite feed system check-out was initiated.

At 11:25 lignite feed table pressurization was changed from flue gas to air. At this stage there was no lignite on the feed tables and no other change in operation. As expected, the gasifier temperature started to increase because of increased stoichiometry but also an unexpected effect was noticed in the form of a black plume from the stack.

At 11:57 lignite feed was started and was working its way through the feed system.

At 12:07 a slight drop in gasifier temperature had been noticed indicating that lignite had reached the gasifier bed; oil flow was reduced to compensate. Lignite and oil gasification conditions, which are summarized in Table C-1, were maintained for almost 2.5 hours before they were discontinued because of frequent blockages in the downcomers to the lignite injectors.

TABLE C-1. COMBINED LIGNITE AND OIL GASIFICATION, FIRST ATTEMPT
(OCTOBER 1, 1980)

From (hr:min)	To (hr:min)	Elapsed minutes	Lignite rate, kg/hr (lb/hr)	Oil rate, kg/hr (lb/hr)	Gasifier temp., °C (°F)
-	12:07	-	0	2,350 (5,185)	911 (1,672)
12:07	12:51	43	600 (1,320)	2,040 (4,500) ^a	907 (1,665)
12:51	13:24	33	980 (2,160)	1,910 (4,200) ^a	909 (1,668)
13:24	14:03	39	1,290 (2,840)	1,760 (3,870)	913 (1,676)
14:03	14:32	29	1,720 (3,800)	1,540 (3,400) ^a	916 (1,681)
14:32	-	-	0	2,210 (4,866)	917 (1,682)

^aEstimated value.

The following observations were made:

- The dark plume from the stack cleared soon after lignite injection started. It was probably caused by a sudden change in the combustion product flow disturbing some of the fines on the boiler tubes.

- Flames in the boiler remained satisfactory.
- At the maximum flow, one-third of the thermal input to the boiler was from lignite.
- Both the regenerator and the ResoxTM plant remained operational.

Completion of Run

At the end of the lignite gasification, conditions were returned to those before lignite injection for a period of 1.5 hours. During that time, bed transfer pulsers were switched from nitrogen to flue gas for a period of 0.5 hour and there was no noticeable difference in performance. At this stage, blockage of flue gas recycle filters had reduced the flue gas availability to the point at which the simultaneous control of gasifier temperature and stoichiometry was becoming difficult and preparations were made to exit the gasification mode. The usual procedure of sulfating the calcium sulfide on the gasifier bed requires sufficient flue gas to fluidize the bed; therefore, that procedure had to be abandoned. Instead, the gasifier and regenerator beds were inerted, slumped, isolated, and allowed to cool by natural convection as follows:

- 16:05: transfer pulsers changed from flue gas operation to nitrogen.
- 16:15: frothing air to oil guns replaced by nitrogen; gasifier temperature 912°C (1673°F).
- 16:20: oil off; reducing air and flue gas by closing off control dampers.
- 16:23: low fluidizing velocity trip; flue gas (B-102) and air (B-101) blowers out of service; closing isolation dampers from blowers. ResoxTM system isolated. The gasifier temperature increased rapidly, peaked at 927°C (1700°F) within a few minutes and then began to fall off slowly; most of the increase was in the front bed temperature while the rear bed temperature hardly changed at all.
- 17:55: gasifier temperature was dropping at the rate of 11°C/hr (20°F/hr); regenerator temperature was 1038°C (1900°F) and dropping.

A controlled shutdown had been achieved with no damaging temperature excursions. Sulfation was attempted by fluidizing with air and slumping, an excessive temperature was achieved. This procedure (tried successfully at ERCA) was attempted on October 2, and was completed successfully on October 3.

Results

Throughout the test run, process and environmental data were monitored and the results are summarized in Table C-2 for a number of selected periods during which conditions were approximately constant. The two values for boiler SO₂ and sulfur removal efficiency in period 4 correspond to before

TABLE C-2. RUN 5 RESULTS: SUMMARY FOR SELECTED PERIODS
(SEPTEMBER-OCTOBER 1980)

Period	1	2	3	4	5	6	7	8	9	10	11
From: day, hr:min	24,12:20	25,16:00	26,09:00	27,00:00	28,10:50	28,20:00	28,23:00	29,09:00	29,21:00	30,10:00	1,12:00
To: day, hr:min	25,16:00	26,08:00	26,20:00	28,10:40	28,19:30	28,22:00	29,07:00	29,21:00	30,10:00	30,10:00	1,16:20
Hours run	24.7	16.0	11.0	34.7	8.7	2.0	8.0	12.0	13.0	27.0	4.3
Boiler load, Mw (gross)	10.3	10.1	10.0	9.9	9.7	10.0	10.5	8.2	8.2	9.1	8.0
Resox operating, % time	0	100	0	74	100	100	100	100	100	100	100
Natural gas, kg/hr	494	758	1052	789	0	472	450 ^a	0	0	54	680
lb/hr	1750	1670	2320	1740	0	1040	1000 ^a	0	0	120	1500
Oil, kg/hr	2300	2232	1987	2304	3470	2354	2400 ^a	2744	2700 ^a	2427	1996
lb/hr	5070	4920	4380	5080	7560	5190	5300 ^a	6050 ^a	6000 ^a	5350	4400 ^b
Limestone, kg/hr	435	231	231	213	0	0	64	209	142	141	0
lb/hr	960	510	510	470	0	0	140	460	310	220	0
Gasifier temp., °C	832	866	871	871	882	899	871	866	871	888	910
°F	1530	1590	1600	1600	1620	1650	1600	1590	1600	1630	1670
Regenerator temp., °C	960	960	966	1021	1066	1088	1071	1077	1082	1066	1082
°F	1760	1760	1770	1870	1950	1990	1960	1970	1980	1950	1980
Boiler SO ₂ , ppm	35	49	82	87/235	550	780	600	710	690	480	560
Boiler O ₂ , volume %	4.3	4.9	5.3	4.9	5.1	5.3	4.1	6.1	5.8	5.5	6.1
Regenerator SO ₂ , %	N.A.	2.3	N.A.	4.5	6.9	N.A.	N.A.	7.6	5.2	5.7	N.A.
Desulfurization efficiency, %	97	95	91	92/78	63	34	59	49	52	66	37 ^a
Boiler efficiency, loss method, %	82	82	83	82	81	82	N.A.	81	N.A.	81	N.A.
Boiler efficiency, in-out, %	93	96	90	94	91	N.A.	N.A.	N.A.	N.A.	N.A.	N.A.

^a Approximate value.

^b Also 612 kg/hr (1350 lb/hr) average lignite.

N.A. = Not available.

and after ResoxTM came on stream and ResoxTM tail gas was fed back to the flue gas recycle fan inlet. There is therefore an apparent anomaly between period 2, when recycling ResoxTM tail gas had no effect on SRE (95 percent), and period 4 in which there was a 14 percent (92 to 78) reduction in SRE on feeding ResoxTM recycle into the gasifier. This difference, in fact, follows from some information that ERCA have obtained on their pilot gasifier, that introduction of steam into the gasifier reduces SRE. During period 2 the regenerator temperature was relatively low and the flow of cooling steam to the regenerator off gas cooler was also relatively low. However, during period 4 (and about the same time as the regenerator off gas was routed to the ResoxTM plant) the regenerator temperature increased, necessitating a flow of extra steam to the off gas cooler, hence to the gasifier via the tail gas, and was probably the root cause of the loss of SRE.

Problems

Problems encountered are similar to those from previous runs:

- blockage of coal injectors
- blockage of cyclone diplegs
- problems with rotary valves
- carryover of fines into the boiler
- problems with initial fluidization; coarse limestone
- regenerator cyclone drain valve timer often did not function.

New problems, which caused more trouble than previously, have been:

- trip-out of startup burners while the coal feed system was in operation
- blockage and damage of filter bags in the flue gas recycle baghouse.

Inspection

A similar pattern of deposits in the boiler as before and some damage to the separating wall between gas passes in the boiler were noticed. There were no problems in the gasifier and regenerator and fewer humps in the pits than have been noticed previously. Cracks in the cyclone drain rotary valve rotors were seen mostly along repair weld lines. Transfer ports were full of unconsolidated material with no sign of blockages. The new thermocouples have survived intact.

There were deposits 10 to 25 cm (4 to 10 in) deep in the gasifier to cyclone ducts.

Flue gas recycle baghouse bags needed replacing and also the bag in the fresh limestone filter receiver.

PART 2: OIL AND LIGNITE GASIFICATION (RUN 6)

Preparations for Run 6

Several major items needed attention before the start of Run 6:

- New bags were installed in the flue gas recycle baghouse S-104. The baghouse cages were in good condition and did not require attention.
- Coal injector needles were shortened further (they had been trimmed back before Run 5) and cut on the slant so that the aperture cross-sections were normal to the gasifier floor.
- Gasifier and regenerator beds were drained and dumped. The gasifier partition wall was repaired.
- The boiler was cleaned and the brickwall between gas passes was repaired.
- Gas regulator for the regenerator startup burner was repositioned to obviate flame outages caused by vibration.
- Failed parts were replaced in the cyclone drain rotary valves.
- Bags were replaced in the S-308 fresh limestone filter receiver.
- The rotor was removed from the B-210 ResoxTM tail gas fan, the casing was modified slightly to reduce gas flow restriction and was sealed to prevent air inleakage.

First Startup Attempt

Initial attempt at fluidizing a charge of fresh limestone was made on November 4, 1980, and was not successful.

On November 5, 1980, after preheating the bed with overhead burners for 6 hours, an attempt to fluidize lifted the front gasifier bed but not the rear bed. At this stage, all the oil gun frothing air was switched to the front oil gun only and lignite feed was initiated to the front bed only. After several hours of trying, the startup attempt was postponed.

On November 6, 1980, heat-soaking bed and refractories.

On November 7, 1980, on attempted fluidization, only the front bed lifted and lignite feed was initiated to the fluidized front bed for a short while because holes had been blown in the front bed and the remainder of the bed slumped.

On November 8, 1980, fluidization attempt once again lifted the front bed, and lignite feed to the front bed was initiated. A sample taken from the gasifier bed showed that the front bed was 60 percent calcined and had a very

poor "dumbbell" distribution; i.e., a mixture of coarse and fine with no mid-fill, which probably accounted for the poor fluidization characteristics. Startup attempts were abandoned when an operator noticed that the "B" startup burner adapter plate was glowing red. The unit was shut down and cooled to allow repairs to the burner refractory to be carried out. During this outage, a weld failure on "B" distributing coal table was repaired.

On November 14, 1980, a further attempt to fluidize and combust was unsuccessful. The current startup attempt was therefore abandoned. Gasifier and regenerator material were drained and dumped, and a supply of fresh limestone with a better size distribution was ordered.

Second Startup Attempt

On November 20, 1980, a new attempt at startup was made with 46 cm (18 in) of freshly delivered limestone in the gasifier bed. This limestone while still fairly coarse, had a much more even size distribution and therefore it was expected to fluidize more easily. Fluidization was initiated at 13:00 with 25,400 kg/hr (56,000 lb/hr) of flue gas, and lignite was injected as soon as the gasifier bed was fluidizing evenly. The gasifier bed temperature increased gradually but unevenly with the rear bed temperature increasing faster: the first oil gun was put in service in the rear bed; at 16:50 this number had been increased to eight. At 21:11, one front oil gun was placed in service. By 21:30 lignite feed and startup burners were taken out of service and the number of front oil guns was increased to five.

Steady Oil Combustion

On November 21, 1980, just after midnight, the gasifier was in steady combustion mode with 494 kg/hr (1,090 lb/hr) of oil fed through 17 oil guns and bed temperature at 927°C (1700°F). Stone feed was initiated at 00:49 at 202 kg/hr (445 lb/hr) to start building up the bed. As conditions stabilized, the limestone flow was increased gradually to the feed system maximum of 794 kg/hr (1,750 lb/hr) and the oil rate was increased by 10 percent to maintain a constant gasifier bed temperature. Since the fresh limestone feed point is located near the regenerator end of the rear bed, the rear bed temperature was lower, at 843°C (1550°F), than the front bed temperature which was 927°C (1700°F).

At 16:50 the bed was slumped for 10 minutes during which the gasifier inspection window was replaced by a solid plug. Combustion was then resumed and continued until the next day. By midnight the gasifier bed level had increased to 66 cm (26 in) and there had been no problems.

On November 22, 1980, early in the morning problems were experienced with transfer of lignite from the main silo to the two surge tanks. The blockages, caused by excessively wet lignite, were cleared and by 05:00 lignite was being transferred, but occasional blockages, mostly at the inlet to the bucket elevator, kept recurring.

By noon, the bed depth had been built up to 86 cm (34 in), lignite surge tanks had been filled, and all systems check out. The limestone rate was reduced to 230 kg/hr (500 lb/hr) and the gasifier operation was switched to the gasification mode.

Steady Oil Gasification

The actual changeover was made at 12:17 using the now well-established method of swapping air for flue gas to reduce the gas duct oxygen to 1 to 2 percent a few minutes before the changeover, then increasing the oil rate sharply from 499 kg/hr to 2,586 kg/hr (1,100 lb/hr to 5,700 lb/hr) to obtain ignition (within a few seconds) and then backing off flue gas to adjust the gasifier stoichiometry. The gasifier settled to steady oil gasification conditions as summarized in Table C-3 and there were no changes or upsets until nearly noon the next day.

TABLE C-3. MEAN OIL GASIFICATION CONDITIONS PRIOR TO LIGNITE/OIL COMBINED GASIFICATION

Item	November 23 rate	November 24 rate
Oil, kg/hr (lb/hr)	2,277 (5,020)	2,087 (4,600)
Plenum air and gas, kg/hr (lb/hr)	14,200 (31,200)	13,200 (29,000)
Plenum O ₂ , volume percent	12.5	12.25
Limestone, kg/hr (lb/hr)	513 ^a (1,130)	769 (1,696)
Temperature, °C (°F)	871 (1,600)	884 (1,624)
Bed level, cm (in)	86.4-96.5 (34-38)	98.0 (38.6)

^aReduced to 340 kg/hr (750 lb/hr) at 7:00 on November 23, 1980.

Combined Oil/Lignite Gasification (First Attempt)

On November 23, 1980, lignite feed check-out and preparations for a changeover were completed by 11:00 and lignite feed to distributing tables was initiated at 12:01 and simultaneously the oil feed rate was reduced from 2,304 to 2,109 kg/hr (5,080 to 4,650 lb/hr). Lignite feed was then increased in stages while oil was backed out, after each lignite increment, to maintain a constant bed temperature of $902 \pm 6^\circ\text{C}$ ($1655 \pm 10^\circ\text{F}$). Limestone feed rate was maintained at 358 kg/hr (790 lb/hr) and flue gas with lignite at 2,180 kg/hr (4,800 lb/hr). The actual steps are detailed in Table C-4.

TABLE C-4. SUMMARY OF COMBINED LIGNITE/OIL GASIFICATION (NOVEMBER 23, 1980)

Time	Oil rate, kg/hr (lb/hr)	Lignite rate, kg/hr (lb/hr)	Plenum air and gas, kg/hr (lb/hr)	Plenum O ₂ , vol. %	Gasifier temperature, °C (°F)
Pre 11:21	2,304 (5,080)	0	13,900 (30,700)	12.55	899 (1650)
11:21-11:50	2,109 (4,650)	726 (1,600)	13,900 (30,700)	12.55	913 (1675)
11:50-12:10	2,028 (4,470)	1,361 (3,000)	14,200 (31,200)	13.48	N.A.
12:10-12:40	1,742 (3,840)	2,288 (5,000)	14,200 (31,200)	13.50	899 (1650)
12:40-13:05	1,633 (3,600)	3,066 (6,760)	14,200 (31,200)	13.50	899 (1650)
13:05-13:17	1,533 (3,380)	3,765 (8,300)	17,700 (32,300)	14.38	899 (1650)
13:17-13:26	1,533 (3,380)	<3,765 ^a (<8,300) ^a	17,700 (32,300)	14.38	increasing
13:26-13:39	1,701 (3,750)	<3,767 ^a (<8,300) ^a	17,700 (32,300)	15.28	927 (1700)
13:39-13:40	1,701 (3,750)	<4,536 ^a (<10,000) ^a	17,700 (32,300)	N.A.	927 (1700)
13:40-14:05	2,077 (4,580)	reducing to 0	14,200 (31,300)	N.A.	falling
After 14:05	2,077 (4,580)	0	14,200 (31,300)	N.A.	904 (1660)

^aActual reading lower than that recorded because of blockage.

N.A. = Not available.

Some rodding was necessary but there were no problems until 13:17 at which point the lignite feed was 3,764 kg/hr (8,300 lb/hr) and the oil rate had been reduced to 1,533 kg/hr (3,380 lb/hr) to maintain a temperature of 899°C (1650°F) when one front and one rear needle were plugged up and gasifier temperature started to rise implying that, although a steady lignite feed rate was indicated, not all of it was reaching the gasifier. At 13:37 the oil rate was increased to 1,701 kg/hr (3,750 lb/hr) to compensate for the temperature increase, and the indicated lignite feed rate was increased to 4,536 kg/hr (10,000 lb/hr); however, problems occurred at this point:

- B distributing table clutch weld failed and that feed table was taken out of service for repairs.
- Lignite had backed up the D table all the way to the weigh feeder so that D table was taken out of service.

Tables B and C were therefore taken out of service, to prevent further problems, and the oil rate was increased initially to 2,040 kg/hr (4,500 lb/hr) for about an hour, during which time the gasifier temperature slowly drifted down from 927°C to 913°C (1700°F to 1675°F). The oil rate was then increased to 2,180 kg/hr (4,800 lb/hr) and conditions were allowed to settle while repairs to the lignite feed system were carried out. An interesting effect was noticed during the lignite/oil gasification period: the indicated gasifier bed depth increased from 96.5 cm to 102 cm (38 in to 40 in) and, during the hour following lignite gasification, it dropped back to 96.5 cm (38 in) implying that lignite char inventory in the gasifier bed was of the order of 5 percent. The only other noteworthy observation was that the Lear Siegler stack opacity increased to 40 percent during lignite/oil gasification and then dropped slowly to a near zero value. Boiler flame was mostly bright and clear, occasionally dusty, throughout the lignite injection period and, until the blockages occurred, there were no operational problems. Steady oil gasification conditions at 2,180 kg/hr (4,800 lb/hr) of oil and a reduced limestone feed rate of 195 kg/hr (430 lb/hr) were maintained until the repairs were completed the next day.

Combined Oil/Lignite Gasification (Second Attempt)

On November 24, 1980, all morning, the oil gasification conditions were steady. At 8:00 the limestone feed rate was increased to 771 kg/hr (1,700 lb/hr) which caused the bed level to increase by 1.5 cm (0.6 in) and the bed temperature to drop by 14°C (25°F) by 9:30 at which point the lignite feed was initiated and, at the same time, the limestone feed rate was reduced to 376 kg/hr (830 lb/hr). Major oil gasification parameters, as recorded during the 9:00 to 9:30 period, are listed in the last column of Table C-3. The lignite feed rate was increased to 5,440 kg/hr (12,000 lb/hr), while the oil rate was decreased to 1,070 kg/hr (2,360 lb/hr) over the next hour; the approximate steps and times are listed in Table C-5 in which the condition after the change is indicated. For example, at 9:36, the oil rate was decreased from 2,087 to 1,905 kg/hr (4,600 to 4,200 lb/hr), simultaneously the lignite rate was increased from zero to 1,016 kg/hr (2,240 lb/hr), plenum air plus gas rate

TABLE C-5. COMBINED OIL/LIGNITE GASIFICATION DURING RUN 6

Time at which change was made	Conditions after the change ^a						
	Oil rate, kg/hr (lb/hr)	Lignite rate, kg/hr (lb/hr)	Air/gas rate, kg/hr (lb/hr)	Gasifier temp., °C (°F)	Limestone rate, kg/hr (lb/hr)	Stack opacity, %	Gasifier bed depth, m (in.)
Initial (9:35)	2,087 (4,600)	0	(0)	890 (1634)	771 (1,700)	4	0.98 (38.6)
9:36	1,905 (4,200)	1,016	(2,240)	890 (1634)	376 (830)	4	0.98 (38.6)
9:46	1,905 (4,200)	1,814 (4,000)	13,400 (29,500)	906 (1662)	376 (830)	13	0.99 (39)
9:55	1,787 (3,940)	2,758 (6,080)	13,500 (29,800)	908 (1666)	376 (830)	26	1.02 (40.1)
10:10	1,479 (3,260)	3,774 (8,320)	13,800 (30,400)	897 (1647)	376 (830)	32	1.02 (40.1)
10:22	1,279 (2,820)	4,645 (10,240)	13,800 (30,400)	889 (1633)	376 (830)	39	1.02 (40.1)
10:35	1,070 (2,360)	5,516 (12,160)	14,000 (30,800)	889 (1633)	0 (0)	42	1.02 (40.1)
10:48	1,107 (2,440)	4,282 (9,440)	14,100 (31,000)	901 (1654)	0 (0)	45	1.05 (41.4)
11:12	1,261 (2,780)	5,806 (12,800)	14,100 (31,000)	899 (1650)	0 (0)	45	1.05 (41.4)
11:22	2,037 (4,500)	0	(0)	899 (1650)	0 (0)	45	1.01 (40.2)
11:46	1,964 (4,330)	0	(0)	911 (1672)	0 (0)	20	1.01 (40.2)

^aLignite feed seal gas rate was 2,286 kg/hr (5,040 lb/hr) throughout.

remained unchanged and limestone rate was reduced from 771 to 376 kg/hr (1,700 to 830 lb/hr). In between 10:35 and 11:12, the lignite feed table A started to back up and then cleared itself while table D backed up and was taken out of service. At 11:12, D table was back in service since the problem appeared to be inadequate flow of wet lignite and not a blockage in the injectors. At 11:22, both B and D tables started backing up and all the lignite feed was taken out of service and oil increased. A few minutes later, the oil rate was reduced to allow the carbon on the bed to burn off. As in the previous run, the gasifier bed increased from an initial value of 98 cm (38.6 in) to a maximum of 105 cm (41.6 in) and dropped back to 98 cm (38.6 in) by 13:00.

Stack opacity also increased appreciably but returned to a low value by the early afternoon. Steady oil gasification with no fresh limestone feed was continued until the next day while repairs were made to the lignite feed system.

On November 25, 1980, oil was gasified steadily all night with no problems. At 8:20 the spent solids cooler drive motor bearings failed and the solids cooler was taken out of service. This was possible because on oil gasification, at a low oil rate with no fresh limestone feed, the quantity of fines passing down the cyclone diplegs was very low. However, it was not possible to attempt any further lignite gasification tests unless the considerable quantity of fine lignite ash and carbonaceous fines generated could be disposed of. Attempts to procure a replacement motor bearing or repair the existing one quickly failed, so an alternative method of disposing of the cyclone fines to the regenerator was tried.

Fines Disposal to Regenerator and Sulfation

At 14:30, 1,409 kg/hr (3,300 lb/hr) of flue gas was admitted to the regenerator plenum, and transfer spargers were placed in pulsing mode. By 14:55, 66 cm (26 in) of gasifier bed material had been transferred to the regenerator and the regenerator temperature was 722°C (1350°F) but started declining so the regenerator overhead burners were put into service. Several attempts were made to feed cyclone fines to the regenerator but problems were encountered with plugging of the water-cooled transfer lines by cyclone fines, probably through condensation of moisture in the flue gas. Solids transfer between the gasifier and regenerator was not very good. This would not have mattered if the regenerator were to be used purely as fluid bed combustor of carbonaceous fines. Fluidization in the regenerator was not even and that could have caused more of a problem. Another and unrelated problem, the low level alarm on the fresh limestone storage tank, forced a decision to abandon this test run, sulfate, and shut down. At 19:21, all oil injection was discontinued and the air isolation damper was closed with flue gas flow to the gasifier plenum at 9,160 kg/hr (20,200 lb/hr) and a gasifier temperature of 899°C (1650°F). Temperature increased to a maximum of 927°C (1700°F) and then declined. Controlled sulfation was completed by 21:23 and oil was injected at 277 kg/hr (611 lb/hr). Oil combustion was continued until 22:15 to burn out the carbon in the ducts and cyclones, and the plant was then shut down and allowed to cool.

Inspection

The gasifier was isolated and not inspected. The regenerator was emptied, cleaned out, and isolated. The boiler was in service on natural gas, although there were some solids deposit, and was not inspected. The flue gas recycle baghouse fines extract auger was jammed. The bags were in good condition. The spent materials tank filter receiver was plugged up.

PART 3: OIL AND LIGNITE GASIFICATION (RUN 7)

Preparation for Run

The spent fines system was cleaned up and motor bearings replaced. The flue gas recycle baghouse was cleaned and the auger repaired. A new limestone shipment was loaded into the fresh storage hopper. Coal distributing tables and the coal feed system were thoroughly checked over.

Startup

A quantity of sulfated bed material measuring 81 cm (32 in) was still in the gasifier from the previous run so no fresh limestone fill was required.

On December 8, 1980, overhead burners were lit and flue gas was blown through the gasifier plenum. At 13:40 flue gas was routed through the coal feed system seal enclosure while maximum flue gas and air were routed through the distributor. At 13:51 the coal feed system was put in operation to feed lignite via tables A, C, and D at 454 kg/hr (1,000 lb/hr) each. The gasifier bed temperature increased steadily and at 15:45 oil guns were put into service with eight guns in the front and eight guns in the rear. Lignite feed to the gasifier was stopped at 16:00 and limestone feed was initiated soon afterwards to build up the bed.

On December 9, 1980, steady combustion was continued all night and morning while equipment check-out was being carried out. At 10:25 gasifier mode of operation was changed from oil combustion to oil gasification with the intention of changing to lignite gasification soon after the first change. In any event, there was a short delay while bearings were changed on the fresh limestone feeder.

Lignite Gasification, First Attempt

At 13:33 while gasifying oil at 2,150 kg/hr (4,730 lb/hr) all the coal needles were rodded and checked for being clear. Lignite feed was initiated at 14:43 at a rate of 908 kg/hr (2,000 lb/hr). Since there were no problems, the lignite feed rate was increased to 5,450 kg/hr (12,000 lb/hr) at 14:09 and the oil rate was reduced to 1,430 kg/hr (3,150 lb/hr). This condition was similar to that at which problems with lignite feed occurred during previous runs; therefore, it was maintained for a while to check for any developments. At 14:45 both cyclone diplegs were found to be glowing orange: thermocouples in

the diplegs were indicating 688°C (1270°F) for the "A" cyclone and 582°C (1080°F) for the "B" cyclone. There were also copious quantities of carbonaceous fines leaking out of the rotary seals at the base of the cyclone diplegs. Lignite feed was therefore stopped and by 15:00 all lignite had cleared out of the feed tables and the gasifier was back on 100 percent oil gasification while damage was assessed. Repairs that were needed required that the plant should be shut down; therefore, at 16:08 all oil was shut off from the gasifier and the bed was sulfated. Sulfation was completed at 17:25 and the gasifier was slumped and isolated.

Lignite Gasification, Second Attempt

On December 10, 1980, both the cyclone drain rotary valves were stripped, seals repacked, and clearances were adjusted. The spent fines filter receiver was cleared out. During the day the gasifier bed was fluidized and the temperature raised via a short period on oil combustion, followed by slump and isolation.

On December 11, 1980, most of the day was spent attempting to route cyclone fines to the regenerator.

At 9:45 the gasifier bed was fluidized and oil combustion initiated. Cyclone fines were routed through to the regenerator and the regenerator itself was started by putting an air flow of 2,220 kg/hr (4,900 lb/hr) to the regenerator plenum and pulsing the transfer spargers at 11:13. By 12:03 regenerator temperature exceeded 650°C (1200°F) and regenerator oil guns were used to boost the regenerator temperature. At 12:30 the cyclone fines line to the regenerator was found blocked. Several unsuccessful attempts later, the cyclone fines were routed back to the solids cooler. A noisy bearing was replaced on the cyclone "A" rotary seal, the temperatures on both the gasifier and the regenerator were pushed up, and both units were slumped and isolated.

On December 12, 1980, in order to make "elbow room," three ResoxTM ash containers were filled with spent fines drained from the spent fines storage vessel.

Oil combustion was initiated in the gasifier bed at 8:44 and the switch to oil gasifying mode was made at 11:15. By 14:00, oil gasification conditions had stabilized, as summarized in Table C-6, and the coal feed system had been checked out. Lignite injection started at 14:48 at 1820 kg/hr (4000 lb/hr) and was rapidly built up to 5450 kg/hr (12,000 lb/hr) while the oil rate was backed down to 860 kg/hr (1900 lb/hr). During the next hour a number of problems occurred, the main problem was lignite backing up or damp lignite not feeding properly onto the feed table but there were no injector blockages. Each table in turn was built up to the maximum rate, then taken out of service to clear and then put back into service. Consequently the total feed rate varied in the range 3633 to 7490 kg/hr (8000 to 16,500 lb/hr); the maximum of 7490 kg/hr (16,500 lb/hr) was maintained for 5 minutes at 15:30 and a rate of 7760 kg/hr (16,000 lb/hr) was established at 15:40 and maintained for 11 minutes. The maximum reliable lignite flow rate was determined for each table as: 680 kg/hr (1500 lb/hr) for tables 'A' and 'D' and 1590 kg/hr (3500 lb/hr) for

tables 'B' and 'C'. At 16:12 the lignite feed rate was established at 4990 kg/hr (11,000 lb/hr) and all oil was switched off. About 3 MW of natural gas was maintained in burner No. 1 and modulated to absorb any sudden changes caused by disturbances in lignite feed rate, also 0.5 MW equivalent of natural gas was fed to the igniters in burners Nos. 3 and 4. These conditions were maintained for 3.5 min and are summarized in Table C-6 alongside the oil data.

No problems were experienced for over 2 hours with gasifier, regenerator or boiler. At 18:30 lignite started backing up on tables 'A' and 'D' and these were taken out of service causing the lignite feed rate to drop to just above 2700 kg/hr (6000 lb/hr). Table 'A' was cleared within a few minutes and was back in service. Table 'D' remained out of service so that the lignite feed rate was raised to 4090 kg/hr (9000 lb/hr) and was maintained at that rate until the end of the run.

End of Run

Once again difficulties were being experienced with the transfer of lignite from the main lignite silo to the surge hoppers. Also, the spent solids hopper was almost full. Lignite feed to the gasifier was therefore discontinued at 19:39 and the bed was sulfated, using flue gas, without transferring back to oil gasification. Sulfation was complete by 20:28, both the gasifier and the regenerator were slumped and isolated. A total of 22.2 Mg (24.5 tons) of lignite had been gasified.

December 13, 1980: The gasifier bed was fluidized with air and, since the temperature started dropping, indicating that the sulfation was complete, oil was injected and combustion was maintained for 1 hour to burn out any overhead deposits.

Inspection

Coal Tables--

Upon inspection of the coal distributing tables after the run, one spring was found failed on both the 'C' and 'D' tables (there are four springs to a table). Also one threaded 'connecting rod' used for spring attachment was found failed on 'B', 'C' and 'D' tables. The pin connection of distributing table drive to the table was found frozen in 'C' and 'D' tables. 'D' distributing table outlet pieces were in some cases found to interfere with downcomer inlet pieces. The 'D' coal rotary air lock feeder feeding its distributing table was found with its rotor pockets considerably plugged with damp lignite.

Cyclone Drains--

Both rotary valves at the bottom of the cyclone drains were warped to the point that severe stress was imposed on the bearings and seal packings while fine adjustment of clearance was no longer possible.

Gasifier--

The partition wall between front and back needed repairs.

TABLE C-6. OPERATING CONDITIONS: OIL ONLY AND LIGNITE ONLY
(DECEMBER 12, 1980)

	Oil Average 14:00-14:40	Lignite Average 16:10-19:39	S.D.
Lignite flow rate, kg/hr	0	4,450	360
Lignite flow rate, lb/hr	0	9,800	800
Oil flow rate, kg/hr	2,320	0	
Oil flow rate, lb/hr	5,120	0	
Air/flue gas to plenum, kg/hr	12,400	10,900	410
Air/flue gas to plenum, lb/hr	27,400	24,000	900
Frothing air, kg/hr	710	710	2
Frothing air, lb/hr	1,570	1,570	4
Flue gas to coal table, kg/hr	2,150	2,310	50
Flue gas to coal table, lb/hr	4,730	5,080	110
Limestone, kg/hr	850	195	14
Limestone, lb/hr	1,870	430	30
Gasifier temperature, °C	983	943	6
Gasifier temperature, °F	1,640	1,730	10
Gasifier depth, in	85.9	92.7	1.2
Gasifier depth, cm	33.8	36.5	0.5
Gasifier plenum O ₂ , vol %	11.85	12.2	0.1
Air flow to regenerator, kg/hr	2,300	2,300	45
Air flow to regenerator, lb/hr	5,000	5,000	100
Regenerator temperature, °C	1,060	1,080	11
Regenerator temperature, °F	1,940	1,990	20

External--

Flue gas recycle blower B-102 bearings were noisy.

Boiler--

There was a nominal build-up of ash and lime fines on the furnace water walls and screen tubes. These were readily tapped off. A veneer of solidified slag was found in the product gas burner throats and on the burner fronts. Samples of boiler deposits were sent for analyses and the results are given in Table C-7.

TABLE C-7. ANALYSIS OF DEPOSITS FROM RUN 7

	Cyclone 'A'	Burner throat	Burner front	Side wall	Rear wall	Super- heater
SiO ₂ wt %	51.5 (58.5)*	35.2	41.1	44.0	5.6	37.7
Al ₂ O ₃ wt %	5.5 (6.3)	4.6	7.5	5.2	1.6	7.1
Fe ₂ O ₃ wt %	2.8 (3.2)	4.8	4.9	4.2	1.7	4.2
CaO wt %	19.5 (22.2)	50.4	41.5	46.9	49.6	35.9
MgO wt %	1.3 (1.5)	1.9	1.9	1.8	1.0	2.0
SO ₃ wt %	4.3 (4.9)	0.6	1.8	0	39.6	12.3

*Note: figures in brackets give the carbon-free analysis.

For comparison, a sample of fines, collected at 16:55 on December 12, 1980 while gasifying lignite, from cyclone 'A' is included in Table C-7, first column. The cyclone sample, in addition to the components specified and 1 to 2 percent of minor components, also contained about 12 percent carbon. Comparison leads to some interesting conclusions:

- Cyclones caught lignite ash fairly successfully.
- The burner throat, burner wall and side wall fines were essentially equal mixtures of calcium oxide and silica, and, where slagging occurred, possibly existed in the form of calcium silicate.
- The rear wall deposit was predominantly calcium sulfate either formed in the flame or in situ.
- The superheater deposits were essentially a 50/50 mixture of the side wall deposit and rear wall deposit.

Several cyclone fines samples, taken during the gasification run were analyzed for carbon and sulfur. Sulfur and carbon stayed fairly constant at about 1.6 percent and 12 percent respectively, as can be seen from Table C-8.

TABLE C-8. GASIFIER CYCLONE MATERIAL ANALYSIS

Location	Taken on: date	At time	Sulfur wt %	Carbon wt %
A	12/12/80	14:30*	5.414	2.5
B	12/12/80	14:30*	3.798	6.3
B	12/12/80	16:30	1.816	12.1
A	12/12/80	16:35	1.701	14.16
B	12/12/80	16:50	1.904	13.82
A	12/12/80	16:55	1.522	14.57
B	12/12/80	17:30	1.753	12.59
A	12/12/80	17:30	1.603	11.74
B	12/12/80	18:45	1.774	9.57
A	12/12/80	18:45	1.759	7.98
B	12/12/80	19:15	1.615	23.88

*Note: taken during oil gasification prior to coal gasification.

Boiler Performance

Throughout the gasification run, boiler load was maintained at approximately 9.5 MW. Product gas from coal provided from 4.0 to 6.5 MW with the balance of heat to the boiler coming from natural gas. Unfortunately, an accurate determination of natural gas flow was not available (the orifice plate for measuring natural gas flow was designed for 9.95 kPa-g (40 in H₂O) while the flow during gasification was from 0.5 to 1.12 kPa-g (2 to 4.5 in H₂O) with fluctuations as great 0.37 kPa (1.5 in H₂O). Therefore, the natural gas flow indications were unreliable. However, since the natural gas was fed to the boiler through two igniters (totaling about 0.5 MW) and one natural gas burner, operating at or less than its 5 MW capacity, the load contribution of natural gas was estimated to be at or less than 5.5 MW. Therefore, based on the burner operating pressure observed during coal gasification (fluctuating between 10.3 and 24.1 kPa-g (1.5 and 3.5 psig)) natural gas provided from 3 to 5.5 MW.

A summary of boiler operating data is shown in Table C-9.

TABLE C-9. BOILER PERFORMANCE

	Average	Highest	Lowest	Standard Deviation
Load, MW	9.52	9.7	9.5	0.07
Feedwater Flow, kg/hr	41,360	43,180	40,900	1,364
Feedwater Flow, lb/hr	(91,000)	(95,000)	(90,000)	(3,000)
Final Superheat Pressure, kPa	4,171	4,206	4,137	16.5
Final Superheat Pressure, psig	(605)	(610)	(600)	(2.4)
Final Superheat Temperature, °C	430	440	424	5.9
Final Superheat Temperature, °F	(806)	(825)	(795)	(10.7)
Feedwater Temperature, °C	170.1	171	168	1.3
Feedwater Temperature, °F	(338.2)	(340)	(335)	(2.4)
Air for Combustion, °C	33.6	35	32	1.4
Air for Combustion, °F	(92.5)	(95)	(90)	(2.6)
Flue Gas Out, °C	216.1	218.3	232	1.2
Flue Gas Out, °F	(421.1)	(425)	(419)	(2.2)
Natural Gas Flow, SCMH	2,359	2,706	1,994	236
Natural Gas Flow, SCFH	(82,800)	(95,000)	(70,000)	(8,300)
Air Flow to Boiler, kg/hr	48,500	52,270	46,360	1,730
Air Flow to Boiler, lb/hr	(106,700)	(115,000)	(102,000)	(3,800)

Throughout the run the product gas flames were bright and definite although somewhat hazy. They were also larger than the oil gasification flame and filled most of the boiler furnace.

Emissions Data

Shown in Table C-10 are averaged values of gaseous and particulate (indicated as opacity) emissions. The oxides of nitrogen emissions are quite low, as they were during oil gasification. Previously obtained emissions data taken during natural gas combustion indicated higher oxides of nitrogen emissions than observed during coal gasification (emissions during natural gas combustion were of the order of 129 ng/J (0.30 lb/10⁶ Btu). Therefore, since natural gas accounted for 35 to 50 percent of the heat input during this test, it is conceivable that if coal derived product gas were the only fuel, oxides of nitrogen emissions could be somewhat lower.

TABLE C-10. GASEOUS EMISSIONS DATA, RUN 7

Location	Oxides of Nitrogen			Sulfur Dioxide		Opacity (%)	Oxygen concentration (Vol %)
	Concentration (ppmv)	ng/J	1b/10 ⁶ Btu	Concentration (ppmv)	Corrected (ppmv)		
Boiler exit	139	119	0.276	257	753	1.31	7.22
Induced draft fan exit	107	91.7	0.213	263	770	1.34	7.20
Stack						78.0	

^aVolume corrected for dilution from natural gas.

^bThis suggests that sulfur was being released.

Sulfur removal efficiency was poor throughout this test as indicated by the sulfur dioxide (SO₂) emissions. Although the SO₂ value measured did not appear excessive, it must be realized that, due to the large quantity of natural gas utilized, the SO₂ was diluted. Therefore, this dilution must be accounted for. The corrected SO₂ value is also shown in the table.

The cause of low SRE was probably the result of several factors. Esso Research (ERCA) had shown that optimal SRE occurred with bed temperatures in the 850 to 900°C (1550 to 1650°F) range, whereas during this test bed temperatures were in the 925 to 950°C (1700 to 1750°F) range. In addition, the gasifier bed had a relatively high sulfur content (11.3 percent) prior to initiation of coal feed. It is also noted that the bed sulfur steadily declined throughout the run (see Table C-11). This might indicate a release of sulfur from the bed instead of a capture, perhaps resulting from the higher bed temperature and stoichiometry (approximately 30 percent). However, the limestone flow maintained during coal gasification should have had a positive effect on SRE, being almost 2.5 times the stoichiometric amount for 100 percent sulfur removal.

TABLE C-11. GASIFIER BED MATERIAL ANALYSIS

Sampled on date	At time	Sulfur wt %	Carbon wt %	Hydrogen wt %
12/10/80	09:30	-	2.5	1.1
12/12/80*	13:00*	11.318	2.3	0.77
12/12/80	16:30	8.035	0.59	0.42
12/12/80	17:10	7.285	2.53	0.71
12/12/80	18:15	5.101	0.42	0.24
12/12/80	19:15	4.212	0.75	0.27

*Note: Taken during oil gasification immediately prior to lignite gasification.

Unfortunately, gasifier conditions that are optimal for high gasification efficiency are not optimal for high SRE. In an integral CAFB-Furnace, where high carbon carryover is tolerable, lower gasifier temperatures may be used and, therefore, higher SRE experienced.

From the outset of coal feed, stack opacity was exceptionally high, ranging from 70 percent to 100 percent. Near the termination of coal feed, opacity showed signs of declining, but not to the point of being satisfactory. Therefore, on any future plant, particulate clean up after the boiler would undoubtedly be necessary (it had previously been shown that

particulate emissions for oil gasification might be considered acceptable). For comparison, standards of performance for new stationary sources for gaseous, liquid and solid fuels are shown in Table C-10 adjacent to the measured values for this test.

PART 4: LIGNITE GASIFICATION (RUN 8)

Preparation

A number of jobs were carried out prior to the test run:

- Coal feed system: springs and eyebolts were replaced (using new ones where possible), all the clutches were flushed, refilled with oil and checked for freedom of movement, replaced weigh cell on 'C' table and recalibrated all tables.
- Gasifier was drained, cleaned, refractory wall was repaired and two drains were fitted.
- Flue gas recycle blower B-102 bearings were shimmed and blower balance was checked.
- Spent material hopper was emptied using a cement mixer truck and removing the material from site as a slurry.
- Lignite was drained from the coal storage silo, air dried and reloaded back into the silo.
- Cyclone drain rotary seals were dismantled, repaired, reassembled and clearances were reset but, since there was some distortion, the rebuild may not have been entirely successful.
- Boiler was taken out of service and cleaned.

Start-up

The gasifier was first fluidized in the morning on February 12, 1981, but since fresh coarse limestone was being used, it took all day to achieve steady oil combustion and, at one point the bed fluidizing velocity was raised to 1.13 m/sec (3.7 ft/sec) to achieve adequate bed mixing. By midnight steady combustion was established with 9 guns in service in the rear and 12 guns in the front, an oil rate of 574 kg/hr (1265 lb/hr) and a high limestone feed rate of 1220 kg/hr (2690 lb/hr) to build up bed.

A number of problems occurred to delay lignite gasification until February 25, 1981. The actual sequence of events is summarized in Table C-12 and since it has no bearing on the lignite gasification, it is not described in more detail.

TABLE C-12. PRE-RUN 8 OIL COMBUSTION AND GASIFICATION
(FEBRUARY 12-24, 1981)

Date (Feb'81)	Gasifier	Regenerator	Notes
12	Start-up	-	Problems with initial fluidization
13	Combustion	-	Building up bed
14	Combustion	-	Water flood in regenerator off-gas cooler and cyclone
15	Combustion (short boost with lignite)	-	Regenerator cyclone drain valve jammed. 'C' table failed
16	Combustion	Drain and clean	Spent fines baghouse plugged: cleared
17	Reheat only	-	Data General computer failed
18	Combustion	Started regenerator	Problems with regenerator oil guns
21	Combustion, change to gasification* at 13:10	Regeneration, but lost regenerator at 16:30	Lump near regenerator to gasifier transfer port
22	Gasification, then sulfation and slump	Drain and rod sequence	-
23	Combustion	Rodding: started 10:00 failed again 14:00	-
24	Combustion	Started transfer at 13:30 temp. up to 760°C (1400°F) using transfer only	Clean spent solids filter and flue gas recycle baghouse

*Note: natural gas flame out as product gas came through but the igniters held and had ignition without a trip-out.

Lignite Gasification

February 25, 1981: at 8:25 the gasifier bed (kept hot and isolated from the previous day) was fluidized and oil combustion initiated. At 9:58 transition to oil gasification was made and transfer of material to the regenerator was initiated by putting the transfer pulsers in service. Transfer between beds was initially poor and did not become established until 11:30 after rodding both the horizontal and vertical slots. As soon as the regenerator was definitely operational, lignite injection was initiated.

In between 11:36 and 13:17 the lignite rate was built up to 5490 kg/hr (12,100 lb/hr). Oil rate was maintained at 2360 kg/hr (5200 lb/hr) until 11:53 it was then reduced gradually to zero at 13:36 at which point the product was based on lignite only. Within minutes of backing out all oil the cyclone diplegs were glowing orange and the regenerator bed temperature rose to 1093°C (2000°F), indicating that both the gasifier bed and cyclone fines had a high carbon content. At 13:45 the gasifier bed temperature was increased to 954°C (1750°F) by increasing the gasifier stoichiometry, to reduce the carbon level both in the gasifier bed and on the fines. At 14:14 some air was backed out of the regenerator and was replaced with recycled flue gas to moderate the regenerator temperature rise. The regenerator temperature stabilized at 1093°C (2000°F) but the cyclone dipleg temperature continued rising and by 14:32 cyclone 'A' dipleg temperature reached 807°C (1485°F) and the 'B' cyclone dipleg reached 739°C (1363°F); the rotary valve on cyclone 'A' dipleg was taken out of service to give it a chance to cool down. Conditions then remained stable for about 1 hour. At 15:29 it was noticed that the spent solids cooler temperature had risen to 138°C (280°F) and an unsuccessful attempt was made to bring it down. At 15:32 lignite was switched off and by 15:37 it had cleared through the feed system and had been replaced by oil, however some damage had occurred already because at 15:35 fires were seen blowing out of the solids cooler. At 15:50 all oil was out of the gasifier and sulfation began. Cooling air to solids cooler was switched off and the solids cooler temperature, which peaked at 193°C (380°F) was dropping. Sulfation was completed by 18:05 and was followed by an hour of oil combustion before slumping the bed.

A total of 16,600 kg (36,520 lb) of lignite was gasified of which 10,632 kg (23,390 lb) was during 1.93 hr of lignite only operation.

Observations

During lignite only operation both the product gas flames were bright and intense but hazy.

There were no recorded problems with the coal feed system or coal injection needles.

Average operating conditions for this run are summarized in Table C-13.

TABLE C-13. RUN 8: LIGNITE GASIFICATION DATA (13:36 TO 15:37 ON FEBRUARY 25, 1981)

Item		Value
Lignite rate	kg/hr	5,490
Lignite rate	lb/hr	12,100
Air/gas to plenum	kg/hr	12,900
Air/gas to plenum	lb/hr	28,400
Plenum O ₂	vol percent	14.18
Recycle gas to coal table	kg/hr	2,100
Recycle gas to coal table	lb/hr	4,700
Limestone feed rate	kg/hr	308
Limestone feed rate	lb/hr	680
Gasifier temperature	°C	943
Gasifier temperature	°F	1,730
Air/gas to regenerator	kg/hr	2,000
Air/gas to regenerator	lb/hr	4,400
Gasifier bed depth	cm	99
Gasifier bed depth	inches	39
Stack opacity (double pass)	percent	35

Inspection

There were no inspection notes available for this test run.

PART 5: BITUMINOUS COAL GASIFICATION (RUN 9)

Preparations for Run

There were no notes available on the preparations made for this test run except for the information that the gasifier bed from the previous run was retained and the gasifier isolated, so that the startup bed was calcined, was quite hot, and contained some lignite ash. No analysis of the bituminous coal was available at this stage except for sulfur which was 2.34 ± 0.08 percent based on 4 analyses of a coned and quartered sample.

Startup and Oil Combustion

March 2, 1981: At 13:35 the hot, calcined bed was fluidized and coal injected using "A" and "B" tables. Oil was injected, in combustion mode, an hour later. Combustion continued until the gasifier was slumped at 18:00 and isolated. During combustion the "B" cyclone drain rotary seal tripped out twice, possibly through overheating.

March 7, 1981: In the evening the gasifier was fluidized and put in combustion mode in preparation for gasification the next day. However, just before midnight a leak was discovered in the "D" feed water heater to the boiler and the gasifier was slumped and isolated to allow repairs to be carried out.

March 8, 1981: Refluidized in the morning for a short warm-up period on combustion on oil but slumped again almost immediately because the frothing air compressor tripped out. An hour-and-a-half later an old site compressor had been tied into the frothing air circuit to ensure an adequate air supply and the reheat combustion was resumed while the frothing air compressor was being repaired.

March 11, 1981: At 00:40 oil combustion was initiated in the gasifier in order to have a long day available for a coal gasification run. Within a few minutes of starting there was a problem with low O₂ alarm on the boiler, caused by sticking vanes on the Forced Draught control. This problem was cleared and combustion resumed but at 1:40 the gasifier rear temperature suddenly dropped. A few minutes later the temperature rose sharply up to a high temperature trip point causing a gasifier shut-down but, fortunately, without tripping out the boiler. On investigation the problem turned out to have been caused by a failed thermocouple and not a process malfunction. The failed thermocouple was disconnected, the gasifier was again refluidized on oil combustion, and preparations for gasification were continued. At 2:04, while switching over oil gun shrouds from air to nitrogen, it was found that there was no air flow at all. The shrouds were put on nitrogen, leaving the investigation of this problem till later. Meanwhile, the faulty thermocouples had been disconnected.

At 3:00, limestone feed was initiated 908 kg/hr (2000 lb/hr) to build up bed, and oil combustion appeared to be settling out except that between 3:30 and 7:30 the rotary seal on gasifier cyclone "B" jammed three times and was cleared and restarted without shutting down the plant. After that oil combustion continued uninterrupted, while the gasifier bed level was built up. Problems were experienced with water condensation and blockages in the spent fines line to the solids cooler. But, by 14:50, that problem had been cleared and air was being backed out of the gasifier in preparation for a changeover to oil gasification.

Oil Gasification

At 15:15 the gasifier operation was changed to the gasification mode at:

- Oil rate = 2,574 kg/hr (5,675 lb/hr)
- Air plus gas rate = 12,740 kg/hr (28,080 lb/hr)
- Plenum O_2 = 12.5 vol percent

Soon after the changeover, attempts were made to start the regenerator. At 16:00 the regenerator appeared to be operational but its temperature reached 910°C (1670°F) and then dropped. Transfer from the gasifier to the regenerator was good, but transfer out of the regenerator was impeded by a heap of unfluidized stone by the transfer slots and possibly a blockage inside the slots. A decision was made, therefore, to slump and isolate the regenerator and proceed with coal gasification without regeneration.

Coal/Oil and Coal Gasification

Coal feed was started to the gasifier at 18:17 and a complete transfer from oil to coal was achieved by 21:12. The schedule of changes is given in Table C-14. During the changeover the gasifier cyclone dip leg temperature increased and by 19:45 "A" cyclone dip leg was glowing red and its temperature was 785°C (1445°F) while "B" dip leg was at 727°C (1340°F). Once on coal only gasification, feeders "B" and "D" behaved well but there were considerable problems with "A" and "C" coal feeders so that the "C" feeder had to be taken out of service twice and "A" feeder once. These interruptions caused the gasifier bed temperature to drift up and therefore, at 22:30 four oil guns were put in service in the front of the gasifier in the area of the "C" coal needles, at a rate of 320 kg/hr (700 lb/hr). By 22:37 all the coal feeders were operational but the oil was left on for 3 more minutes to help bring the temperature down. At 23:23 the "D" feeder was lost for a while and four oil guns, in the vicinity of "D" table needles, were put in service at 340 kg/hr (740 lb/hr) to lower the gasifier temperature.

At 23:35 "D" table was back on stream and all oil was again backed out of the gasifier.

March 12, 1981: At 00:15 "C" table tripped out, then backed up a few minutes later, then tripped and was taken out of service and, since the gasifier temperature started drifting up again, oil injection was started on

TABLE C-14. RUN 9: SCHEDULE OF COAL AND OIL RATE CHANGES

Time	Oil		Notes	Coal		Notes
	kg/hr	(lb/hr)		kg/hr	(lb/hr)	
18:00	2,450	(5,410)	On oil only	0		---
18:15	2,450	(5,410)	S	940	(2,080)	Coal on
18:22	2,030	(4,480)	D	940	(2,080)	S
18:27	2,030	(4,480)	S	2,030	(4,480)	I
18:31	1,490	(3,280)	D	2,030	(4,480)	S
18:43	1,490	(3,280)	S	2,900	(6,400)	I
18:52	1,190	(2,620)	D	2,900	(6,400)	S
19:01	1,490	(3,280)	I	2,140	(4,720)	"C" table out
19:12	1,190	(2,620)	D	2,900	(6,400)	"C" table in
19:40	1,510	(3,330)	I	1,670	(3,680)	"A" and "C" out
19:50	1,090	(2,400)	D	2,900	(6,400)	"A" and "C" in
20:12	1,090	(2,400)	R	2,900	(6,400)	S
21:12	0	0	Oil off	2,900	(6,400)	S
21:28	0	0	---	2,100	(4,640)	"C" table out
21:40	0	0	---	2,900	(6,400)	"C" table in
22:20	0	0	---	2,100	(4,640)	"C" table out
22:25	0	0	---	2,900	(6,400)	"C" table in
22:30	320	(700)	I	2,100	(4,640)	"C" table out
22:40	0	0	D	2,900	(6,400)	"C" table in
23:23	0	0	---	2,250	(4,960)	"D" table out
23:26	340	(740)	I	2,250	(4,960)	S
23:30	340	(740)	S	2,900	(6,400)	"D" table in
23:35	0	0	D	2,900	(6,400)	S
24:00	0	0	---	2,090	(4,600)	Between midnight and 00:32 "C" table tripped out 4 times
00:32	360	(800)	I	2,210	(4,880)	"C" table out of service
01:00	360	(800)	S	1,420	(3,120)	"D" table out
01:03	360	(800)	S	2,210	(4,880)	"D" table in
01:14	360	(800)	S	1,420	(3,120)	"D" table out of service
01:18	710	(1,570)	I	1,420	(3,120)	S
01:37	0	0	Oil off	0	0	Coal off

Notes: S = steady
D = decrease
I = increase
R = ramping down from

four guns in the vicinity of "C" coal feed needles at a rate of 360 kg/hr (800 lb/hr). At 1:00 "D" table was taken out of service for 3 minutes in order to clear it and there was nothing in it to clean because damp coal had bogged down in the bottom of the surge hopper and was not descending. At this stage the conditions seemed somewhat adverse and little useful data was being generated, therefore both oil and coal ceased at 01:37 and the bed was sulfated. Mechanical troubles have not ceased even at this stage because, during sulfation another rear bed thermocouple failed.

Results

Coal gasification data for the first bituminous coal run are summarized in Table C-15. These data are approximate because of the very unsteady nature of coal gasification during this period. A total of 17,980 kg (39,630 lb) of coal were gasified.

Inspection

Gasifier--one thermowell failed and all others were severely pitted. Several coal needles were also badly pitted. Burner throat refractory on "B" starter burner was cracked. An area 1.5 by 1.5 m (5 ft by 5 ft) of gasifier hot face refractory needed removing and replacing. "C" table end plate had rotated through an angle and prevented normal operation.

New cyclone drain rotary valve internals were ordered.

Boiler--nominal dust deposits, no boiler pipes, no fused deposits.

PART 6: COAL GASIFICATION (RUN 10)

Preparation for Run

The boiler tubes were sand blasted and the boiler cleaned in preparation for the statutory insurance inspection.

The gasifier refractory was repaired; two replacement thermocouples were purchased from Omega Engineering and installed; and the burner throat was repaired.

A 35.6 cm (14 in) scale of fine bed material was removed from the gasifier offtake ducts.

New internals were installed in the Ducon rotary seals at the bottom of the cyclone dip legs.

Startup Attempts

Between May 13 and May 22, 1981 six unsuccessful attempts were made to start the gasifier. A sieve analysis eventually indicated that the root cause of the startup problems was that the limestone was much too coarse.

TABLE C-15. SUMMARY OF THE MAJOR PROCESS CONDITIONS
(FIRST BITUMINOUS COAL GASIFICATION RUN)

Coal rate kg/hr	2,900
Coal rate (lb/hr)	(6,400)
Air/gas rate kg/hr	13,340
Air/gas rate (lb/hr)	(27,200)
Gasifier plenum O ₂ vol percent	13.03
Gasifier temperature °C	980-1070
Gasifier temperature (°F)	(1800-1950)
Boiler load, MW	8.7
Natural gas contribution, MW	4.8
Product gas contribution, MW	3.9
Gasifier carbon utilization, percent	69
Sulfur removal efficiency, percent	9-48
SO ₂ emission ng/J	486-856
SO ₂ emission (lb/10 ⁶ Btu)	(1.13-1.99)
NO _x emission ng/J	86.1
NO _x emission (lb/10 ⁶ Btu)	(0.2)
Opacity (Lear Siegler), percent	20

Consequently the gasifier was emptied and a new supply of finer limestone was blown into the gasifier bed directly from the supply truck. Coarse limestone from site storage was used for topping up.

Apart from the fluidization problems, the frothing air compressor tripped out on May 14, and, on inspection, the lubrication system was found to be gummed.

Startup

May 29, 1981: The gasifier and regenerator were preheated with the overhead burner. At 14:45 the gasifier was fluidized and 680 kg/hr (1500 lb/hr) of coal was fed to the gasifier. While the temperature was increasing slowly a number of problems occurred:

- The spent fines line to the solids cooler plugged up.
- The "B" coal feed table appeared to feed more coal than was indicated.
- The cyclone drain leg Ducon rotary seals tripped out several times and periodically needed resetting and restarting.
- A great deal of fine dust was being carried over into the boiler.

A gasifier trip was experienced at 17:22 for an unidentified reason. The gasifier was refluidized at 17:37 and the enforced heat soak caused a rapid temperature rise in the gasifier so that at 18:00 oil guns were put in service. The limestone feed failed soon afterwards, causing an appreciable drop in bed level over the next few hours until the bed was slumped to conserve stone. Between May 30 and June 7, the bed was refluidized daily on combustion mode but could not be switched to gasification mode because a number of small mechanical problems occurred:

- The boiler feed water regulator needed repairing.
- The Zone 4 bed thermocouple burned out.
- The cyclone down rotary feeder jammed requiring the plant to be shut down and the rotary clearances to be readjusted.
- Power was lost to the "A" fines injector control panel.
- The recycle gas fines extraction auger was blinded with fines and needed manual clean out.
- On June 5 the operation was changed to oil gasification but two thermocouples were lost in the gasifier bed. The bed was sulfated and returned to the oil combustion mode.
- On June 6 the nitrogen onsite ran out.

On June 7, 1981 the bed was restarted on combustion mode and changed to oil gasification mode at 11:00. The regenerator was not in service at this stage of operation.

Oil gasification was continued throughout the night.

Coal Gasification (First Attempt)

On June 8, 1981 at 01:11, coal feed to the gasifier was initiated at 910 kg/hr (2,000 lb/hr). However, "D" table was not feeding because of a blocked rotary feeder and soon afterwards there were problems with coal feed from the "C" table. At 2:09, the cyclone "A" dip leg rotary seal tripped out and a few minutes later the "B" dip leg rotary seal jammed. At 2:28, gasification was transferred back to oil and at 3:48 the oil was switched off and the bed was sulfated. Sulfation was completed by 4:33, then oil combustion was continued for a short while to burn off any carbonaceous deposits, and to bring up the temperature. The bed was then slumped to carry out repairs.

Coal Gasification (Main Run)

On June 9, 1981, oil combustion was initiated at 8:21 followed by a change to oil gasification at 10:00. The coal feed system was started at 11:31 but was stopped at 12:13 to carry out a number of jobs:

- The filter on the outlet from the solids cooler was cleaned.
- A plugged line from the spent fines cooler was cleaned.
- The vibrating conveyor in the coal transfer system was repaired.

These tasks were carried out without shutting down while gasifying oil at 2,500 kg/hr (5,520 lb/hr).

At 14:20, coal feed was initiated again and the usual slow transition from 100 percent oil gasification to 100 percent coal gasification was completed by 15:36 with the coal rate steady at 3,990 kg/hr (8,800 lb/hr). Coal gasification then settled down with the only problems being a regular blockage of the filter on the spent solids cooler outlet, and the "B" cyclone drip leg rotary seal, which tripped out early in the coal gasification run and stayed out. This gasification period lasted 5.5 hours ending in a high temperature trip. The salient conditions of the run are summarized in Table C-16. Throughout this run boiler flames stayed clear and bright, similar to oil gasification flame but clearer.

The high temperature trip at 21:15 was caused by bridging in the downcomer to the "C" and "D" (i.e., front) feeders, causing a total loss of feed on the gasifier front at about 12:12 and a temperature rise of 110°C (200°F) in 3 minutes. The gasifier was restarted immediately in the gasification mode with "C" and "D" feeders at a slightly reduced rate. At 21:25 the "C" and "D" feeders bridged again and oil was injected at 950 kg/hr (2,100 lb/hr) to control the gasifier temperature. At 21:40, the "C" and "D"

TABLE C-16. RUN 10: COAL GASIFICATION CONDITIONS SUMMARY

	Test 1	Test 2
Day	June 9, 1981	June 9-10, 1981
Time	15:36-21:12	22:42-5:40
Coal rate kg/hr	3,990	4,220
Coal rate (lb/hr)	(8,800)	(9,300)
Air/gas rate kg/hr	12,660	12,700
Air/gas rate (lb/hr)	(27,900)	(28,100)
Plenum O ₂ , vol percent	13.62	13.10
Gasifier temperature, °C	1013	982
Gasifier temperature, (°F)	(1855)	(1800)
Gasifier bed depth, cm	101.6	101.6
Gasifier bed depth, (in)	(40)	(40)
Stack SO ₂ , vol ppm	490	500
Stack NO _x , vol ppm	100	115
Stack O ₂ , vol percent	3.9	4.1
Cyclone carbon, wt percent	70-75	70-75
Gasifier bed carbon, wt percent	0.5-2.8	0.5-2.8

tables were brought back on stream but the "B" table was taken out of service to clear three blocked injector needles so that the coal rate was down to 2,860 kg/hr (6,302 lb/hr). All tables were on at 22:09 but the "A" and "B" tables were out again almost immediately causing a coal feed reduction to 1,590 kg/hr (3,500 lb/hr). All four tables were in service again at 22:40 at 4,100 kg/hr (9,040 lb/hr) and all the oil was backed out at 22:42. At 23:30, an attempt was made to rod out cyclone "A" which was running at a suspiciously low temperature. The dip leg was found to be blocked, and, after it was cleared and carbonaceous fines drained out, the rotary seal temperature increased rapidly to 760°C (1400°F) whereupon it jammed. At 4:45 the front gasifier bed temperature increased sharply to 1093°C (2000°F) but it was identified as an instrument fault. Throughout this run the main coal system was in use (i.e., crusher, grinder, bucket elevator and Redler conveyor) and coal was transferred successfully, with only minor problems to the two surge tanks. By 5:40, the plant was out of coal and transition was made to oil gasification after a run of 7 hours. A total of 55,974 kg of coal (61.7 tons) had been gasified during the two runs of 5.5 and 7 hours, plus 1 hour of oil to coal transition and 1 hour of coal feed problems. At 5:50 the oil feed to the gasifier was stopped to initiate sulfation, which was completed at 7:20. This was followed by a short combustion period on oil.

PART 7: BITUMINOUS COAL GASIFICATION (RUN 11)

Preparations for Run

The gasifier bed material was reused from Run 10.

A fresh supply of West Virginia bituminous coal was delivered to the site from Valley Builders storage. Analysis of the as received coal is given in Table C-17.

A natural gas sample, taken at La Palma, was analyzed with the results detailed in Table C-18.

Startup and Combustion

On June 22, 1981, startup posed no problems because the 101.6 cm (40 in) of gasifier bed had been calcined and fluidized before, and was still warm. Oil combustion was continued for 10 hours while the usual startup problems, such as blockage of the fines line to the solids cooler and blocked pressure tappings, were attended to. The usual equipment check out was made.

The regenerator (which had been drained, cleaned and had transfer slots cleared) was not put in service at this point. Instead a decision was made to start it during oil gasification. This approach had been tried before at La Palma with little success but, since it had been used successfully at ERCA, it was decided to try it again.

TABLE C-17. WEST VIRGINIA COAL ANALYSIS

Component	Wt percent
Carbon	75.15
Hydrogen	4.82
Oxygen	4.93
Nitrogen	1.12
Sulfur	2.02
Ash	7.83
H ₂ O	4.13
Higher heating value, MJ/kg (Btu/lb)	31.2 (13,407)

TABLE C-18. NATURAL GAS ANALYSIS

Component	Wt percent
CH ₄	93.282
C ₂ H ₆	3.828
C ₃ H ₈	0.944
C ₄ H ₁₀	0.561
C ₅ H ₁₂	0.190
C ₆ H ₁₄₊	0.387
H ₂	0.328
CO ₂	0.482
Mean molecular wt	17.587
Higher heating value, MJ/m ³ (Btu/scf)	39.5 (1059.5)

Oil Gasification

Transition from oil combustion to gasification was made at 11:00 with no problems. Oil gasification continued at 400 kg/hr (5,300 lb/hr) for an hour and then was increased to 2,950 kg/hr (6,500 lb/hr). It held at that rate until the initiation of coal feed at 14:33.

During this time the regenerator was being preheated with the overhead burners and preparations were made to start it. However, at 11:45 the Gemco drain valves at the bottom of the regenerator cyclone dip leg started tripping out and needed attention. Regenerator startup was therefore postponed until later.

Coal Gasification

Transition from 100 percent oil gasification to 100 percent coal gasification was made between 14:33 and 15:35, finishing with a coal flow of 3,629 kg/hr (8,000 lb/hr). However, at 15:45, the "B" coal feed was tripped out because of blocked needles. Oil feed to this area kept the gasifier temperature from rising excessively. Problems with the "B" table continued intermittently until 17:38 at which point steady coal gasification with no oil injection was established. Steady coal gasification continued until just after midnight when "B" coal feed table caused problems again and oil was injected to control gasifier temperature. At 1:25 the "B" table was put out of service, the feed rate on "A", "C" and "D" tables was increased to compensate, and oil feed was switched off. At 2:35, on June 23, 1981, coal feed table "D" was taken out of service in an attempt to balance the coal requirement of the two sides of the gasifier. The feed rate on tables "A" and "C" was increased to about 1,800 kg/hr (4,000 lb/hr) each, however desulfurization deteriorated markedly and there were indications of problems with "C" table.

Consequently, at 3:15 all coal was taken off the gasifier and sulfation was initiated. Sulfation was finished at 3:50 and oil combustion was initiated for a short period before shutting down.

Results

The total duration of oil/coal and coal gasification was 12.75 hours divided as follows:

- (a) 3.5 hours oil and coal at various rates
- (b) 6.9 hours coal only, all four tables in use
- (c) 1.58 hours coal only, A, C, and D tables in use
- (d) 0.67 hours coal only, A and C tables only.

The average data for the coal-only period (9.2 hours) is summarized in Table C-19.

TABLE C-19. SUMMARY OF COAL ONLY GASIFICATION DATA

Item	
Boiler load MW	10.7
Product gas contribution, MW	4.8
Coal feed, kg/hr (lb/hr)	3,300 (7,300)
Gasifier bed temperature, °C (°F)	1010 (1850)
Carbon conversion, percent	67
Sulfur removal efficiency, percent	71-44
Product gas CO, vol percent	12.6
Product gas CO ₂ , vol percent	12.0
Total coal gasified, Mg (long tons)	39.9 (40)

Operation under the three coal gasification regimes: (b) fully balanced, (c) unbalanced, and (d) partly balanced was different although the coal feed rate was approximately the same in the three cases. Data for the three cases is given in Table C-20. Data for condition (a), partial oil and coal gasification, is not included because it was for a varying condition and because oil was switched on for short periods during upset conditions when equipment was malfunctioning.

TABLE C-20. DATA FOR DIFFERENT COAL GASIFICATION PERIODS (RUN 11)

Item	(b)	(c)	(d)
Period designation			
Duration hours	1.9	1.6	0.7
Coal tables in use	A,B,C,D	A,B,D	A,C
Coal feed rate, kg/hr (lb/hr)	3,433 (7,570)	3,090 (6,810)	3,250 (7,170)
Plenum gas flow, kg/hr (lb/hr)	11,800 (26,000)	11,800 (26,100)	11,200 (24,600)
Plenum O ₂ , vol % dry	13.0	12.8	11.8
Coal table gas flow, kg/hr (lb/hr)	3,060 (6,750)	3,100 (6,840)	3,100 (6,840)
Coal table gas flow, O ₂ , vol % dry	6.8	6.1	6.2
Frothing air flow, kg/hr (lb/hr)	500 (1,100)	500 (1,100)	500 (1,100)
Limestone rate, kg/hr (lb/hr)	310 (690)	290 (650)	260 (570)
Gasifier bed temperature, °C (°F)	1,001 (1,833)	1,021 (1,870)	1,027 (1,880)
Boiler emissions, SO ₂ , vol ppm	420	830	920
Boiler emissions, NO _x , vol ppm	85	83	70
Boiler emissions, CO, vol ppm	290	370	1600
Boiler O ₂ , vol % dry	6.7	4.6	5.5
Boiler CO, vol % dry	11.1	11.1	10.2
Opacity, %	33	41	45
Sulfur removal efficiency, %	70	--	44
Carbon utilization, %	74	--	--

APPENDIX D

The following tables present the results of the laboratory tests conducted by GCA/Technology Division on samples of flue gas collected during oil gasification Run 5. The flue gas samples were collected using the source assessment sampling system (SASS). The SASS train collects particulate matter and inorganic trace species in three cyclones with cut sizes of 10 μm , 3 μm , and 1 μm , a particle filter, and two types of impingers (isopropyl alcohol, and hydrogen peroxide). The first four tables present inductively coupled argon plasma spectrometry (ICAP) results of the peroxide and alcohol impingers and their blanks. The samples were collected during two separate SASS train runs, each of approximately 8 hours, and combined to provide a sufficient quantity of sample for analysis. The fifth table presents the atomic absorption (AA) results of the alcohol impinger blank, the combined alcohol impingers, the peroxide impinger blank, and the combined peroxide impinger and condensate. The sixth table presents the spark source mass spectroscopy (SSMS) results. The first column are the results of the <1 μm particles and the particle filter combined, the second column are the <3 μm and 10 μm particles combined, the third column is the total, and the fourth column is the total divided by the volume of air sampled or the total mass emission rate.

Project 2-045-012GCA Control No. 9904 & 9905 &
9927 & 9928

DATA REPORT SHEET

Metals

Sample I.D. SBI-3-03-HI & SBI-3-03-CD-AO &
SBI-3A-03-HI & SBI-3A-03-CD-AO Report Date 5/6/81
 Sample Matrix Peroxide Impinger Number of Replicates 3
 Analyst(s) A. Cutter Checked By D. Bause

Element	Instrument	Concentration ($\mu\text{g/ml}$)	Emission ($\mu\text{g/ml}$)
Aluminum	Jarrell Ash 855 ICAP	N/A	N/A
Antimony	"	< 0.015	< 1.4
Arsenic	"	< 0.032	< 2.9
Barium	"	< 0.0006	< 0.05
Beryllium	"	< 0.0004	< 0.04
Boron	"	N/A	N/A
Cadmium	"	< 0.007	0.63
Chromium	"	0.120	11.0
Cobalt	"	0.047	4.3
Copper	"	0.249	23.0
Iron	"	N/A	N/A
Lead	"	< 0.016	< 1.4
Magnesium	"	N/A	N/A
Manganese	"	0.012	1.1
Mercury	Perkin-Elmer 460 AAS	N/A	Cold Vapor Method
Molybdenum	Jarrell Ash 855 ICAP	N/A	N/A
Nickel	"	0.492	45.0
Selenium	"	< 0.022	< 2.0
Silver	"	N/A	N/A
Thallium	"	N/A	N/A
Vanadium	"	0.017	1.5
Titanium	"	0.003	0.27
Zinc	"	0.662	60.0

N/A Not Analyzed

Project 2-045-012GCA Control No. 9926

DATA REPORT SHEET

Metals

Sample I.D. _____ Report Date 5/6/81
Sample Matrix Peroxide Blank Number of Replicates _____
Analyst(s) A. Cutter Checked By D. Bause

Element	Instrument	Concentration (µg/ml)	Remarks
Aluminum	Jarrell Ash 855 ICAP	N/A	
Antimony	"	ND*	
Arsenic	"	ND	
Barium	"	ND	
Beryllium	"	ND	
Boron	"	N/A	
Cadmium	"	ND	
Chromium	"	ND	
Cobalt	"	ND	
Copper	"	ND	
Iron	"	N/A	
Lead	"	ND	
Magnesium	"	N/A	
Manganese	"	ND	
Mercury	Perkin-Elmer 460 AAS	N/A	Cold Vapor Method
Molybdenum	Jarrell Ash 855 ICAP	N/A	
Nickel	"	ND	
Selenium	"	0.034	
Silver	"	N/A	
Thallium	"	N/A	
Vanadium	"	ND	
Titanium	"	ND	
Zinc	"	0.002	

* Not Detected
N/A Not Analyzed

Project 2-045-012GCA Control No. 9902 & 9903 &
9924 & 9925

DATA REPORT SHEET

Metals

SBI-3-03-AI2 & SBI-3-03-AI3 &

Sample I.D. SBI-3A-03-AI2 & SBI-3A-03-AI3 Report Date 5/6/81Sample Matrix APS Impinger Solution Number of Replicates 3Analyst(s) A. Cutter Checked By D. Bause

Element	Instrument	Concentration ($\mu\text{g/ml}$)	Emission $\mu\text{g/m}$
Aluminum	Jarrell Ash 855 ICAP	N/A	N/A
Antimony	"	< 0.015	< 0.63
Arsenic	"	0.130	5.5
Barium	"	< 0.0006	< 0.03
Beryllium	"	< 0.0004	< 0.017
Boron	"	N/A	N/A
Cadmium	"	0.014	0.59
Chromium	"	0.199	8.4
Cobalt	"	< 0.003	< 0.13
Copper	"	0.103	4.3
Iron	"	N/A	N/A
Lead	"	0.060	2.5
Magnesium	"	N/A	N/A
Manganese	"	< 0.001	< 0.04
Mercury	Perkin-Elmer 460 AAS	N/A	Cold Vapor Method
Molybdenum	Jarrell Ash 855 ICAP	N/A	N/A
Nickel	"	0.151	6.4
Selenium	"	0.592	25.0
Silver	"	N/A	N/A
Thallium	"	N/A	N/A
Vanadium	"	< 0.0054	< 0.23
Titanium	"	< 0.0012	< 0.05
Zinc	"	0.728	31.0

N/A Not Analyzed

Project 2-045-012GCA Control No. 9901 & 9923

DATA REPORT SHEET

Metals

Sample I.D. SBI-3-03-AIB & SBI-3A-03-AIB Report Date 5/6/81
Sample Matrix APS Impinger Blank Number of Replicates 3
Analyst(s) A. Cutter Checked By D. Bause

Element	Instrument	Concentration (ug/ml)	Remarks
Aluminum	Jarrell Ash 855 ICAP	N/A	
Antimony	"	ND *	
Arsenic	"	ND	
Barium	"	ND	
Beryllium	"	ND	
Boron	"	N/A	
Cadmium	"	ND	
Chromium	"	ND	
Cobalt	"	ND	
Copper	"	ND	
Iron	"	N/A	
Lead	"	ND	
Magnesium	"	N/A	
Manganese	"	0.001	
Mercury	Perkin-Elmer 460 AAS	N/A	Cold Vapor Method
Molybdenum	Jarrell Ash 855 ICAP	N/A	
Nickel	"	ND	
Selenium	"	ND	
Silver	"	N/A	
Thallium	"	N/A	
Vanadium	"	ND	
Titanium	"	ND	
Zinc	"	0.007	

* Not Detected
N/A Not Analyzed

RESULTS OF ATOMIC ABSORPTION SPECTROMETRY

GCA Contract No. 2-045-012

GCA Control No.	Sample Identification	Li [*]		Na [*]		Sr [*]	
		ug/ml	Emission ₃ ug/m	ug/ml	Emission ₃ ug/m	ug/ml	Emission ₃ ug/m
9923 & 9901	SBI-3A-03-AIB SBI-3-03-AIB	< 0.10		3.0		< 0.5	
9924/25 & 9902/03	SBI-3A-03-AI ^a SBI-3-03-A ^a	< 0.10	< 4.2	3.0	130	< 0.5	< 21
9926	SBI-3A-03-CHB	< 0.10		0.4		< 0.5	
9904/05 & 9927/28	SBI-3-03-CH ^b SBI-3A-03-CH	< 0.10	< 7.1	1.6	110	< 0.5	< 36

* Blank corrected.

^a 2nd and 3rd impingers.

^b 1st impinger and condensate.

RESULTS OF SPARK SOURCE MASS SPECTROSCOPY

CONTRACT NO. : 2-045-012

AIR VOLUME (DSCM) : 55.45

ELEMENT	FF-100H ANS 3+3A (MG)	3+100H-C ANS 3+3A (MG)	TOTAL 5425 (MG)	EMISSION FOUND (MG/DSCM)
U	0.003	0.005	0.008	0.001
TH	0.005	0.005	0.011	0.001
BI	0.001	0.003	0.004	0.001
FB	0.35	0.045	0.40	0.007
TL	0.002	0.003	0.005	0.001
Hg	NR	NR	NR	NR
AU	0.002	0.004	0.006	0.001
PT	0.007	0.012	0.019	0.001
IR	0.004	0.006	0.010	0.001
OS	0.004	0.007	0.011	0.001
RE	0.003	0.004	0.007	0.001
W	0.003	0.003	0.006	0.001
TA	NR	NR	NR	NR
HF	0.004	0.014	0.018	0.001
LU	0.001	0.003	0.004	0.001
YB	0.001	0.002	0.004	0.001
TM	0.001	0.001	0.002	0.001
ER	0.003	0.004	0.007	0.001
HQ	0.001	0.001	0.002	0.001
DY	0.002	0.011	0.011	0.001
TB	0.001	0.001	0.002	0.001
GD	0.002	0.002	0.004	0.001
EU	0.001	0.002	0.003	0.001
SM	0.003	0.005	0.008	0.001
ND	0.003	0.067	0.087	0.002
FK	0.001	0.028	0.028	0.001
CE	0.004	0.073	0.077	0.001
LA	0.002	0.31	0.31	0.006
BA	0.20	0.22	0.42	0.008
CS	0.008	0.001	0.009	0.001
I	0.002	0.003	0.003	0.001
TE	0.002	0.003	0.005	0.001
SB	0.010	0.001	0.010	0.001
SN	0.013	0.001	0.014	0.001
IN	IS	IS	IS	IS
CD	0.020	0.003	0.023	0.001
AG	NR	NR	NR	NR
PD	0.002	0.004	0.006	0.001
RH	0.001	0.001	0.002	0.001

RESULTS OF SPARK SOURCE MASS SPECTROSCOPY (con't)

ELEMENT	APPROXIMATE RMS 3+34 a (MG)	3+10UM-C RMS 3+34 b (MG)	TOTAL BASE (MG)	EMISSION FOUND (MG/DSCM)
CU	0.002	0.003	0.006	0.001
MO	0.005	0.010	0.016	0.001
NB	0.001	0.003	0.003	0.001
ZR	0.004	0.003	0.007	0.001
Y	0.001	0.023	0.023	0.001
SR	0.05	1.6	1.6	0.032
RS	0.006	0.010	0.016	0.001
BR	0.003	0.007	0.010	0.001
BE	0.001	0.002	0.003	0.001
AS	0.030	0.028	0.058	0.001
GE	0.001	0.001	0.002	0.001
GA	0.019	0.017	0.037	0.001
ZN	7.0	1.0	8.0	0.15
CU	0.090	0.018	0.11	0.002
NI	1.5	0.80	2.3	0.042
CO	0.021	0.14	0.16	0.003
FE	2.7	12	15	0.27
MN	0.074	1.2	1.2	0.022
CR	0.031	0.20	0.24	0.004
V	3.1	3.5	6.6	0.12
TI	0.079	0.33	0.41	0.007
SC	0.015	0.015	0.029	0.001
LA	39	MC	ee	ee
F	MC	1.0	ee	ee
CL	NR	0.42	0.42	0.008
S	MC	MC	ee	ee
P	0.20	1.3	1.3	0.024
SI	21	39	59	1.1
AL	39	28	68	1.2
MG	18	2.4	2.4	0.043
NA	MC	3.6	ee	TO
F	0.38	0.80	0.80	0.33
O	NR	NR	NR	ee
N	NR	NR	NR	0.014
C	NR	NR	NR	NR
B	9.4	0.095	0.095	NR
BE	0.001	0.001	0.002	TO
LI	0.085	0.001	0.085	0.001
				0.002

^a Particle filter and 1 μ m cyclone catch.

^b 3 and 10 μ m cyclone catches.

